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Evaluation of the Limestone Dual Alkali Prototype System Plant Scholz System Design and Program Plan

Thyssen-CEA Environmental Systems, Inc.
New York

Prepared for

Industrial Environmental Research Lab.
Research Triangle Park, NC

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16. ABSTRACT The report describes the (Phase I) design of both a limestone dual alkali system at Gulf Power Co.'s Scholz steam plant and a related test program. The limestone dual alkali process will be tested at an existing 20 MWe prototype facility at the Scholz plant. The intent of the project is not only to demonstrate the technical feasibility of the process at a prototype level but also to supply technical and cost information related to the implementation of the process at a full commercial scale. The project is being carried out in three phases: Phase I, project planning; Phase II, procurement, recommissioning, and construction; and Phase III, startup, teating, decommissioning, and data analysis. Recommissioning costs at Scholz were about $800,000. Capital and operating costs (in 1979 dollars) projected for a new 500 MW limestone dual alkali system based on the prototype experience are estimated to be $95/kW and 4.5 mills/kWh, respectively. The process was developed by Combustion Equipment Associates, Inc. and Arthur D. Little, Inc.; CEA's Air Pollution Division was sold to Thyssen-CEA Environmental Systems, Inc. in late 1980.

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Coal	Stationary Sources	21D
Combustion	Dual Alkali Process	21B
Desulfurization		07A,07D
Limestone		08G
Scrubbers		13I

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EPA-600/7-81-141a
August 1981

EVALUATION OF THE LIMESTONE DUAL ALKALI PROTOTYPE SYSTEM

PLANT SCHOLZ: SYSTEM DESIGN

AND PROGRAM PLAN

by

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ABSTRACT

The limestone dual alkali process developed by Combustion Equipment Associates, Inc. (Ownership of CEA's Air Pollution Division was sold to Thyssen-CEA Environmental Systems, Inc. in late 1980.) and Arthur D. Little, Inc. will be tested at an existing 20 MWe prototype facility at Gulf Power Company's Scholz Steam Plant. The intent of this project is not only to demonstrate the technical feasibility of the process at a prototype level but also to supply technical and cost information related to the implementation of the process at a full, commercial scale.

The project will be carried out in three phases: Phase I - Project Planning; Phase II - Procurement, Recommissioning, and Construction; and Phase III - Startup, Testing, Decommissioning, and Data Analysis.

This report covers the work in Phase I of the project including the: design and description of the limestone dual alkali system at Scholz; test program itself; recommissioning and construction costs; and projected costs for a full-scale 500 Mw utility system.

CONTENTS

	Page No.
Abstract	ii
List of Figures	iv
List of Tables	v
Applicable Conversion Factors English to Metric Units	vii
Acknowledgements	viii
I. SUMMARY	1
A. Overall Purpose and Scope	1
B. CEA/ADL Limestone Dual Alkali System	2
C. Test Program	2
D. Economic Considerations	3
II. INTRODUCTION	4
A. Background	4
B. Project Objectives	5
C. Scope of Work	5
D. Schedule	6
III. THE LIMESTONE PROTOTYPE SYSTEM AT SCHOLZ	10
A. Limestone Dual Alkali Technology	10
B. Description of the Scholz Plant	11
C. System Design	15

CONTENTS (Continued)

		Page No.
IV.	TEST PROGRAM	32
	A. Schedule	32
	B. Test Program Phases	33
	C. Data Requirements	37
	D. Data Gathering	37
	E. Data Reduction and Analysis	43
V.	ECONOMIC CONSIDERATIONS	52
	A. Costs of Modifying the Scholz Prototype System	52
	B. Generalized Cost Estimates for a New 500 Mw System	52

FIGURES

Figure No.		Page No.
II-1	Project Organization	7
II-2	Limestone Dual Alkali Prototype Test Program Schedule	8
IV-1	Sampling Locations	40

DRAWINGS

Process Flow Diagram 8101-1-1, Rev 6	12

TABLES

Table No.		Page No.
III-1	Limestone Dual Alkali System - Design Basis	17
III-2	Basis for Process Material Balances - Normal Operation	18
III-3	Material Balance - Normal Operation	19
III-4	Basis for Process Material Balances - SO_2 Injection	23
III-5	Material Balance - SO_2 Injection	24
III-6	Assumptions for the Water and System Volume Balances	28
III-7	System Volume Balance and Surge Capacity Requirements - Normal Operation	29
III-8	System Volume Balance and Surge Capacity Requirements - SO_2 Injection	31
IV-1	Prototype Test Program - Projected Operating Conditions	34
IV-2	Process Parameters and Other Conditions Which Affect the Performance of the Limestone-Based Dual Alkali System	35
IV-3	System Performance Evaluation - Data Requirements	36
IV-4	Data Log Format	39
IV-5	Sampling Requirements	41
IV-6	Analyses of Process Samples	42
IV-7	Auxiliary Data	44
IV-8	Procedures for Reducing and Analyzing the Limestone Dual Alkali Data	45

TABLES (Continued)

Table No.		Page No.
V-1	Projected Costs for Modification of the Dual Alkali System	53
V-2	Design Basis for the Generalized Dual Alkali System	54
V-3	Estimated Capital Investment for a 500 Mw Limestone Dual Alkali System	56
V-4	Estimated Cost of Process Equipment and Materials for a 500 Mw Limestone Dual Alkali System	57
V-5	Estimated Annual Operating Costs for a 500 Mw Limestone Dual Alkali System	59

APPLICABLE CONVERSION FACTORS

ENGLISH TO METRIC UNITS

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in	2.54 centimeters
1 in^2	6.452 $centimeters^2$
1 in^3	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal	3.7853 liters
1 Btu	252 calories

ACKNOWLEDGEMENTS

This report was prepared by Arthur D. Little, Inc.; however the information and data contained in the report represent the work of many individuals from several organizations who have been involved in this project. The principal participating organizations are the U.S. Environmental Protection Agency, Thyssen-CEA Environmental Systems, Inc., Gulf Power Company, Inc., Southern Company Services, Inc., and Arthur D. Little, Inc.

We would like to acknowledge the support of the Gulf Power Company and Southern Company Services, which as co-sponsors of the project with the EPA and Thyssen-CEA Environmental Systems, Inc., have contributed equipment, facilities, and the time of their staff. Randall Rush and Lamar Larrimore of Southern Company Services have been instrumental in the overall coordination of the project and have made important technical contributions. The cooperation and support of George Layman and William Lyford of the Gulf Power Company are also greatly appreciated. Their assistance throughout planning and design phases has been an important element in the overall success of the project.

In addition, we would like to acknowledge the efforts and contributions of the EPA. Norman Kaplan, the EPA Project Officer for this prototype test program, and Michael Maxwell, the Director of Emissions/Effluent Technology at EPA's Industrial Environmental Research Laboratory, have provided invaluable guidance and support throughout the development and implementation of this project.

I. SUMMARY

A. OVERALL PURPOSE AND SCOPE

The overall purpose of this program is to evaluate the performance at a prototype scale of the limestone dual alkali process developed by Combustion Equipment Associates, Inc. (CEA)[1] and Arthur D. Little, Inc. (ADL). The intent of this project is not only to demonstrate the technical feasibility of the process at a prototype level but also to supply technical and cost information related to the implementation of the process at a full, commercial scale.

The program will be carried out at an existing 20 Mw prototype dual alkali facility at Gulf Power Company's (GPC) Scholz Steam Plant. CEA and ADL will modify and convert the 20 Mw lime-based dual alkali system for operation with limestone. Following its conversion, the system will undergo a two month startup and a six month program of testing. The project will be carried out in three phases[2]:

- Phase I - project planning;
- Phase II - procurement, recommissioning, and construction; and
- Phase III - startup, testing, decommissioning, and data analysis.

The project is being carried under the sponsorship of EPA, GPC, Southern Company[3], and CEA. The principal cost of the project will be shared by EPA

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[1]On October 20, 1980, Combustion Equipment Associates, Inc. filed a petition for financial reorganization pursuant to Chapter 11 of the U.S. Bankruptcy Code. Thyssen-CEA Environmental Systems, Inc. (TESI) bought ownership of the CEA Air Pollution Division and began administering this EPA contract in early December 1980. TESI assumed financial responsibilities retroactive to October 21, 1980. Although the EPA contract, subject of this project, has been formally assigned to TESI in March 1981, CEA will still be mentioned in this report in connection to work or contributions actually made by CEA.

[2]Phases I and II were completed in early 1980; startup, however, was delayed until August 1980. The delay in starting the system was due to the need to coordinate the process operation with a waste disposal test program sponsored by the Electric Power Research Institute (EPRI). CEA's declaration of bankruptcy in October caused further delays. The actual testing of the system began in February 1981.

[3]Southern Company is an electric utility holding company operating in the Southeast; it is the parent company of Alabama Power Company, Georgia Power Company, Gulf Power Company, Mississippi Power Company, and Southern Company Services, Inc.

and CEA. GPC will provide for use the site and basic services. Southern Company Services' engineering staff will assist in coordinating the project and provide support in developing the test program plan.

B. CEA/ADL LIMESTONE DUAL ALKALI SYSTEM

The 20 Mw prototype system, located at the Scholz Steam Plant (Sneads, Florida), was designed and has been previously operated as a lime-based dual alkali process. The system will be recommissioned and modified to operate with limestone, in a closed-loop mode and in accordance with the following criteria:

- Operation in a "concentrated" alkali mode (>1.0 M active alkali);
- Operation with minimum sodium consumption;
- Operation with minimum calcium consumption (capability for 95+% limestone utilization);
- Capability for removal of more than 95% of the incoming sulfur dioxide;
- Production of a solid waste filter cake containing a minimum of 55% insoluble solids (calcium sulfite/sulfate); and
- Reliable operation with normal variation of:
  - flue gas composition from the power plant,
  - boiler load,
  - electrostatic precipitator operation, and
  - limestone quality.

C. TEST PROGRAM

The test program will begin with the startup of the system and a breakin testing period. The system will then undergo a period of formal testing to characterize its operation. The test program will evaluate the following aspects of the limestone dual alkali system performance:

- SO_2 removal capabilities.
- Consumption of raw materials--soda ash, limestone, and water--not only as a measure of the efficiency of the system in utilizing these materials in the removal of SO_2 but also as a measure of the ability of the system to maintain a closed-loop operation.
- Consumption of energy--electrical power, and fuel for flue gas reheat.
- Level of oxidation experienced by the system as well as the ability of the system to tolerate such level of oxidation.

- Waste cake properties--dewatering, handling, and washability.
- Particulate matter removal (during periods of planned precipitation turndown) by the system and the effect of ash on process chemistry.
- Sulfate emissions in the processed flue gas.

Originally, the test program was to be conducted in two parts. During the first part, lasting approximately two months, a baseline for the operation of the system would be established. The second part, lasting approximately four months, would consist of a series of short variational tests, during which the response and performance of the system to charges in operating conditions would be evaluated. Budgetary constraints, due to actual costs exceeding those budgeted, delays due to the need to coordinate with the EPRI program, and additional delays caused by CEA's declaration of bankruptcy, have forced the reduction of the testing period to only two months.

D. ECONOMIC CONSIDERATIONS

The cost of recommissioning and modifying the dual alkali system at Scholz (Phases I and II of this project) was originally estimated at $613,500. Of this amount, EPA was to cover a total of $366,300 and CEA was to contribute the remaining $247,200. (The actual CEA contribution exceeded $400,000.)

The capital investment for a new 500 Mw limestone dual alkali system is estimated at $47.4 million. The direct costs--process equipment, other materials, erection costs, services, utilities, and miscellaneous--amount to $25.6 million. The indirect costs--engineering design, construction field expense, contractor's fee, and contingency--are estimated to be $11.5 million. The interest during construction and allowance for startup and modifications amount to $8.2 million. Finally, the cost of land, allowance for spare parts, and working capital are estimated at $2.1 million. This investment represents a cost of $94.8/kw. The total operating costs (including capital charges) for this limestone dual alkali system are estimated at $15.7 million per year, representing an annualized cost of 4.5 mills/kwh or $191 per ton of SO_2 removed.

II. INTRODUCTION

A. BACKGROUND

Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) developed, over the past few years, a dual alkali process for removing SO_2 from flue gas generated in coal-fired utility boilers. The CEA/ADL dual alkali process is based on the absorption of SO_2 in an alkaline sodium solution, followed by regeneration of the absorbing solution by reaction with a second alkali, calcium. These reactions generate insoluble calcium-sulfur salts which are discharged from the system as a moist cake.

The CEA/ADL process presents two significant advantages over conventional direct lime or limestone scrubbing technology. First, it uses a clear liquor, rather than a slurry, for scrubbing the flue gas. Second, the regeneration of the scrubbing solution and precipitation of waste solids take place outside the absorber. Thus, the potential for scaling and plugging in the absorber is minimized; the formation of solids with good dewatering properties is more easily controlled; and high SO_2 removal efficiencies (>90%) are easily achieved with alkaline sodium scrubbing solutions by simple manipulation of the scrubber operating pH.

The CEA/ADL dual alkali process has been tested extensively at laboratory, pilot plant, and prototype levels using lime (calcium hydroxide) as the source of calcium for the regeneration reactions[1]. The performance of the process in these test programs prompted the federal Environmental Protection Agency (EPA) to select the CEA/ADL dual alkali process for a demonstration plant. This lime-based demonstration plant has been installed on a 300 Mw boiler at Louisville Gas and Electric's Cane Run Station. The system is currently undergoing a one-year test program.

Although the lime-based system is technically and economically viable, a source of calcium cheaper than lime would increase the economic attractiveness of the process. Limestone (calcium carbonate) was recognized early on as a potential, cheaper, source of calcium for the dual alkali process. Extensive testing of the use of limestone was undertaken by ADL at laboratory and pilot plant levels under funding both by EPA[1,2] and CEA. Successful pilot operations using limestone were finally achieved in 1977. In an extended run using Fredonia Limestone, SO_2 removal efficiencies greater than 90% were attained with limestone utilization of 99+% and

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[1] LaMantia, C.R., et al., "Final Report: Dual Alkali Test and Evaluation Program," Volume I-III, EPA-600/7-77-050a,b,c, May 1977.

[2] Oberholtzer, J.E., et al., "Laboratory Study of Limestone Regeneration in Dual Alkali Systems," EPA-600/7-77-074, July 1977.

waste filter cake containing 55 to 60% insoluble solids. In other testing performed with different limestones, the reactivity and efficiency of utilization of limestone was found to depend importantly on its particle size distribution--the smaller the particles, the better the performance of the limestone.

The successful performance of the system during the laboratory and pilot plant test programs revealed the potential of the limestone-based dual alkali process for full-scale applications. This project is aimed at confirming this potential as well as providing needed information for its full-scale application.

B. PROJECT OBJECTIVES

The objectives of this program are to demonstrate the technical and economic feasibility of the CEA/ADL limestone dual alkali technology for application in flue gas desulfurization (FGD) of high sulfur, coal-fired boilers.

Under this program, the existing lime dual alkali system at Gulf Power Company's (GPC) Scholz Steam Plant will be modified and converted for the exclusive use of limestone as the regenerating material.

After modification and conversion to a limestone-based process, the system will undergo a test program to evaluate its performance with respect to:

- Capabilities for removal of SO_2.
- Raw materials and energy requirements.
- Quality of the waste material generated.
- Reliability and ease of operation of the system.

Furthermore, capital and operating costs will be estimated for full-scale utility applications of the limestone dual alkali FGD technology.

Parallel to this EPA program, the Electric Power Research Institute (EPRI) is sponsoring a program to study and evaluate the landfill disposal of the waste cake generated by the limestone dual alkali system. Prior to its disposal, the FGD waste cake will be treated with fly ash and lime. The EPRI test program will consist of: monitoring the waste disposal system to characterize its operation; analyzing the physical and chemical properties of the treated waste during and after disposal operations; and establishing the potential for revegetation of the waste disposal area. The natural interdependancy between the FGD and the waste disposal system calls for close coordination between the EPA and EPRI programs.

C. SCOPE OF WORK

The scope of work for this project includes the modifications and conversion of the existing system to a limestone system, as well as a six month period of testing.

CEA[3], as the prime contractor to EPA, has the overall responsibility for all aspects of the project. ADL, as a subcontractor to CEA, will provide process engineering support to CEA in designing the modifications to the system, developing the test plan, and assisting CEA in the startup of the process and in coordinating and supervising the test program. ADL is also responsible for the preparation of all reports required under the EPA/CEA contract. The management organization for carrying out the limestone dual alkali project is shown in Figure II-1.

The work is divided into three phases:

- Phase I - program planning;
- Phase II - procurement and construction; and
- Phase III - test program.

This report covers the work performed under Phase I of the project. During this phase, CEA and ADL were to:

- Design the necessary modifications to the existing prototype facilities;
- Develop an overall program schedule;
- Develop a detailed test plan; and
- Estimate capital and operating costs for a full-scale limestone dual alkali system.

D. SCHEDULE

The original schedule for the limestone dual alkali prototype test program is given in Figure II-2. Work on Phases I and II commenced immediately after EPA awarded the contract to CEA in October 1978. Phase I covered a period of seventeen months, ending in February 1980. Phase II, modification of the system, was completed in March 1980. The system was started up in August 1980. In the interim, the system was operated mechanically in order to prevent mechanical deterioration of the equipment. The delay was caused by equipment delivery problems encountered in the installation of the EPRI waste disposal system. As was previously indicated, EPRI is sponsoring a parallel test program to study the landfill disposal of waste generated by the limestone dual alkali process.

[3]And later, Thyssen-CEA Environmental Systems, Inc. See footnote 1 on page 1.

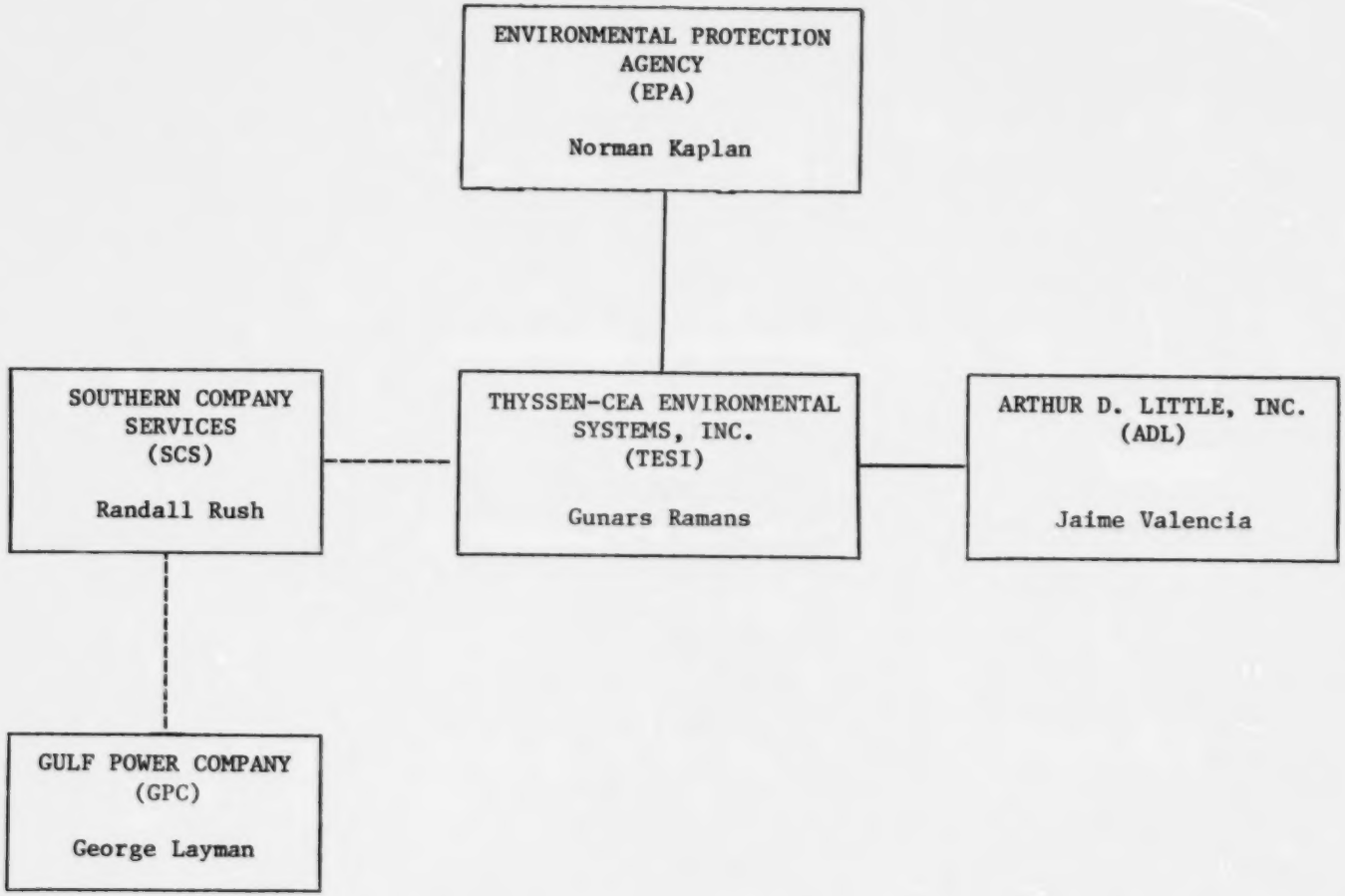
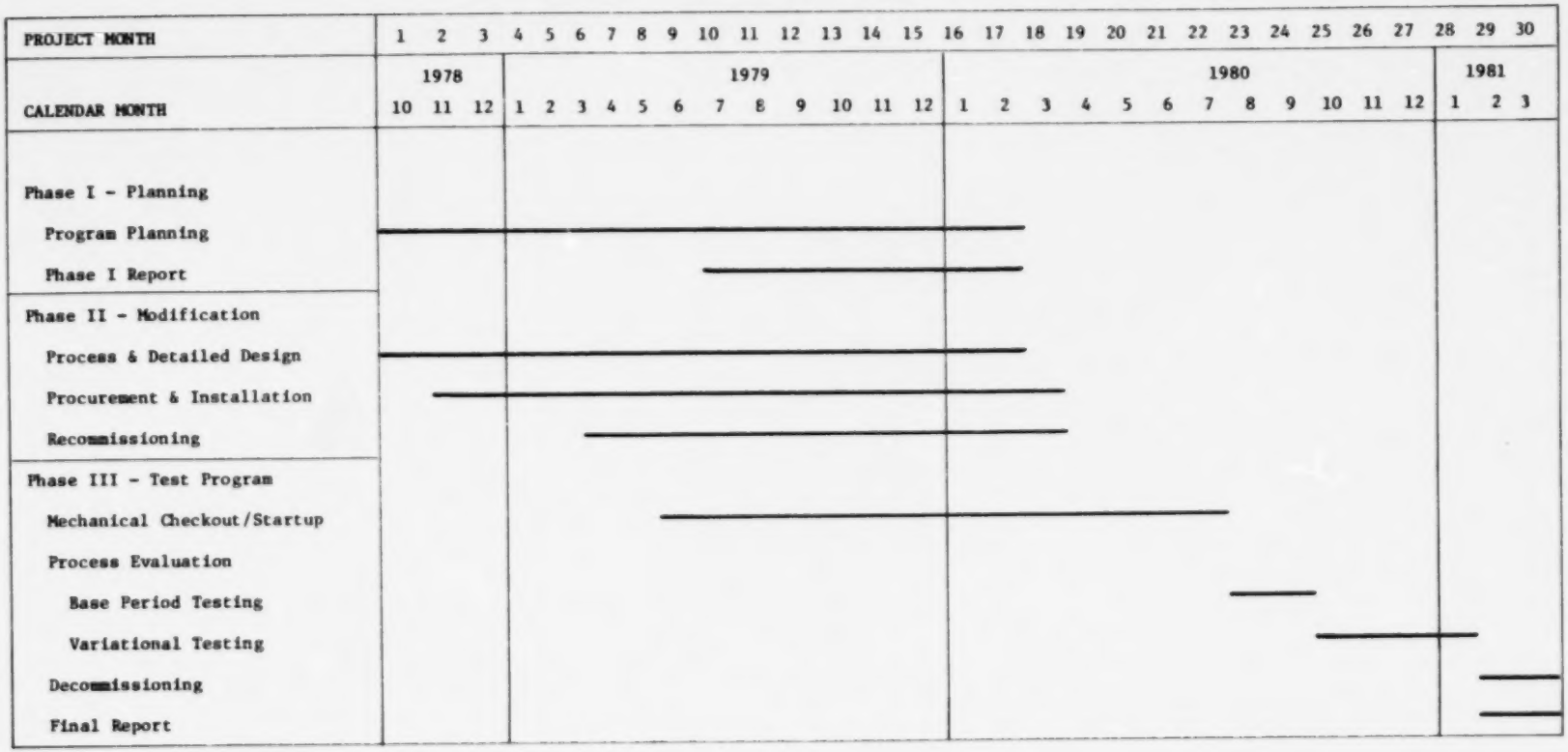


FIGURE II-1: PROJECT ORGANIZATION

FIGURE II-2

LIMESTONE DUAL ALKALI PROTOTYPE TEST PROGRAM SCHEDULE


PROJECT MONTH
1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 21 22 23 24 25 26 27 28 29 30
1978
1979
1980
1981
CALENDAR MONTH
10 11 12 1 2 3 4 5 6 7 8 9 10 11 12 1 2 3 4 5 6 7 8 9 10 11 12 1 2 3
Phase I - Planning
Program Planning
Phase I Report
Phase II - Modification
Process & Detailed Design
Procurement & Installation
Recommissioning
Phase III - Test Program
Mechanical Checkout/Startup
Process Evaluation
Base Period Testing
Variational Testing
Decommissioning
Final Report

The EPA test program of the FGD system was originally scheduled to last six months exclusive of startup procedures and it was to be conducted in two periods. During the first period, approximately two months, a baseline for the operations of the system was to be established. The second period, approximately four months, was to consist of variational testing of the system. However, because of budgetary constraints, due to actual costs exceeding those budgeted, delays due to the need to coordinate with the EPRI program, and additional delays caused by CEA's declaration of bankruptcy, the testing period has been reduced from six to two months. These two months will be dedicated, primarily, to baseline testing, which began in February 1981.

III. THE LIMESTONE PROTOTYPE SYSTEM AT SCHOLZ

A. LIMESTONE DUAL ALKALI TECHNOLOGY

The CEA/ADL limestone dual alkali technology incorporates SO_2 absorption; absorbent regeneration; waste solids dewatering; and raw materials storage and feed preparation.

Absorption of SO_2 is accomplished by contacting the flue gas with a sodium sulfite/bisulfite solution. The sodium sulfite reacts with the SO_2 producing additional sodium bisulfite according to the following overall reaction:

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

During absorption, and to a lesser extent throughout the remainder of the system, some sulfite is oxidized to sulfate:

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4$$

The level of oxidation experienced during absorption is generally a function of the scrubber configuration, oxygen content of the flue gas, and the scrubber operating temperature. At excess oxygen concentrations normally encountered in utility power plant operations, the level of oxidation is expected to be on the order of 5 to 10% of the sulfur dioxide removed (for medium and high sulfur coal applications).

The spent scrubbing solution is reacted with limestone to regenerate the absorbent. The reaction precipitates mixed calcium sulfite and sulfate solids, resulting in a slurry containing up to 5 wt.% insoluble solids. The regeneration process involves basically the following overall reactions:

$$2NaHSO_3 + CaCO_3 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2\ H_2O\downarrow + 1/2\ H_2O + CO_2\uparrow$$

$$2NaHSO_3 + Na_2SO_4 + CaCO_3 \rightarrow 2Na_2SO_3 + CaSO_4 \cdot 1/2\ H_2O\downarrow + 1/2\ H_2O + CO_2\uparrow$$

Following regeneration, the insoluble calcium-sulfur salts are separated from the regenerated liquor. After settling in a thickener, the insoluble salts are vacuum filtered prior to being discharged as a moist cake. The clarified liquor from the thickener is returned to the scrubber, closing the process liquor loop.

The filter cake is washed to recover most of the soluble sodium salts in the mother liquor. A portion of the sodium salts remains occluded in the cake and is lost. Thus, some sodium carbonate is added to the system to make up for these sodium losses. Typically, the Na_2CO_3 makeup amounts to less than 5 mole % of the SO_2 removed.

The amount of limestone added to the system for absorbent regeneration is reduced slightly due to the soda ash makeup. Under normal conditions, the limestone feed stoichiometry will be slightly less than one mole of $CaCO_3$ per mole of SO_2 removed.

B. DESCRIPTION OF THE SCHOLZ PLANT

The Scholz Steam Plant, located in Sneads, Florida, is the property of the Gulf Power Company. The facilities in this plant, as they relate to the present project, are best described in three sections: the boiler, the limestone dual alkali system, and the waste disposal facilities.

1. Boiler

The dual alkali system will take flue gas from Unit No. 1, a 40 Mw nominal capacity (47 Mw peak capacity) Babcock and Wilcox pulverized-coal-fired power boiler. The flue gas from the boiler passes through a high efficiency, sectionalized electrostatic precipitator designed to remove up to 99.5% of the particulate matter. Part of the gas discharged from the precipitator is directed to the limestone dual alkali system, which is designed to handle a flue gas flow rate equivalent to a 20 Mw boiler load. The remaining gas is sent directly to the stack.

2. Limestone Dual Alkali System

The limestone dual alkali system at Scholz is a modification of a lime dual alkali system previously installed and operated at the plant for a test program sponsored by Southern Company and jointly funded with EPA and EPRI. The modified dual alkali system consists of four sections as discussed previously: absorption; regeneration; waste solids dewatering; and raw materials storage and feed preparation. A process flow diagram for the limestone dual alkali system for Scholz is given in CEA drawing 8101-1-1[1].

a. Absorption

The absorption system consists of a variable-throat, plump-bob type venturi scrubber followed by an absorption tower. The venturi can be used for SO_2 absorption and/or particulate removal, and can be operated on a separate liquor loop from the absorption tower or in series with the absorber liquor loop. The absorption tower can be operated as a tray tower, spray tower, or a deentrainment separator.

A sulfur dioxide injection system is provided to increase the SO_2 concentration in the flue gas entering the dual alkali system by 500 ppm or more. The SO_2 is injected upstream of the booster fan before the gas enters the venturi scrubber.

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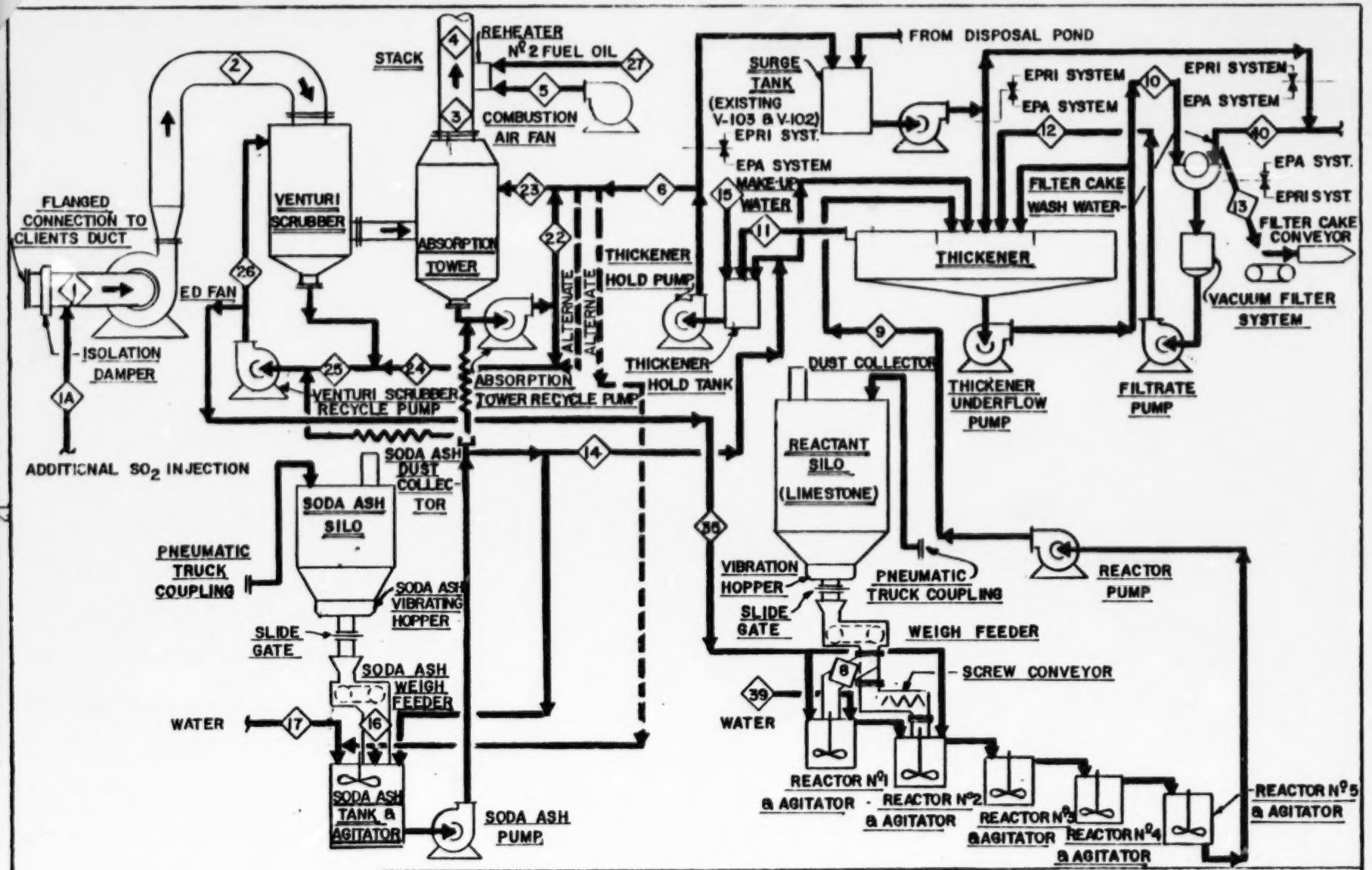
[1] A set of large size plant elevation and layout diagrams and process and instrumentation diagrams may be obtained from the IERL Project Officer at the Environmental Protection Agency, Research Triangle Park, North Carolina. Because of too many details, these drawings cannot be reduced; and, therefore, are not included here.



COMBUSTION EQUIPMENT ASSOCIATES, INCORPORATED NEW YORK, N.Y.	PROCESS FLOW DIAGRAM	LIMESTONE DUAL ALKALI FGD PROCESS GULF POWER COMPANY PLANT SCHOLZ-UNIT NO: 1
	8101-1-1	R6

A booster fan directs the SO_2-rich flue gas exiting the electrostatic precipitator into the scrubber, where it flows downward over the wetted approach section and into the high velocity venturi throat. Recycled scrubbing liquor also enters the top of the scrubber through tangential pipe inlets and through a number of vertical bull nozzles equally spaced around the center of the venturi. After passing through the throat, the saturated gas and scrubbing liquor flow downward through the internal downcomer. The gas makes a 180° upward turn and passes to the absorption tower. The liquor is collected at the bottom of the scrubber in the internal recycle tank.

After being quenched and saturated with water, the partially desulfurized gas enters the bottom of the absorption tower. This tower was designed to operate either as a spray tower (with one to three sets of sprays) or as a tray tower (with one to four trays). For operation in the limestone dual alkali mode, the tower will be operated with two trays. The bottom tray is equipped with a spray underneath to wet the bottom side of the tray. Gas passes upward through the trays and then through a deentrainment separator. The deentrainment separator is operated dry, without mist eliminator wash water.

Scrubbed flue gas leaving the tower is finally reheated by the injection of the hot gas from an oil-fired reheater before discharging through the stack on top of the absorber.

The scrubbing solution, which is fed to the top tray of the absorber, flows countercurrent to the gas through the tray system. This counterflow permits high SO_2 transfer from gas to liquid phase. The liquor is collected at the bottom of the absorber in the internal recycle tank, which serves as the supply for the liquor sprays beneath the bottom tray. Provision is also made to recycle liquor from the internal tank to mix with fresh feed should it be required to maintain liquid loading at low boiler loads.

A bleed from the absorber liquor recycle line is sent forward on recycle tank level control to the venturi recirculation loop where it is used to quench the gas and provide additional SO_2 removal. A bleed stream of spent liquor is drawn from the venturi recycle line and is fed to the absorbent regeneration system at a rate controlled by the liquid level in the venturi internal tank.

b. Absorbent Regeneration

Spent scrubber solution from the recirculation line is bled (on level control) to a multi-stage reactor system which can be operated in a number of configurations. The reactor system consists of five separate tanks: the existing primary reactor followed by four new equal-sized tanks in series with a total design holdup time of approximately 100 minutes. The new tanks are identical in design, each being a cylindrical, baffled vessel with a center-mounted agitator.

The new tanks are located adjacent to the existing reactor system for ease in interfacing with the limestone feed system and the existing primary reactor tank, and to allow for greater flexibility in the reactor system operation. The bleed from the scrubber system is piped such that it can be fed either to the existing first stage reactor or to the first of the four new reactor tanks. Similarly, the modified limestone silo feed chute has the flexibility of feeding dry limestone to either of these two tanks. Thus, the modified reactor system is capable of being operated as either four or five tanks in series with dry limestone feed, or as four tanks in series with slurried limestone feed prepared in the existing first stage reactor. The first four tanks operate on overflow, while the last tank is on level control so that the effluent can be pumped to the thickener.

c. Waste Solids Dewatering

Slurry from the reactor system is pumped to the center well of the existing slurry thickener. Clear liquor overflow from the thickener is collected in the existing thickener hold tank which acts as surge capacity for the absorbent liquor feed to the scrubber system. Process makeup water is added to this tank to make up for the total water lost due to evaporation and moisture in the cake.

The thickened slurry is fed to the existing rotary drum vacuum filter. The underflow slurry is pumped through a recirculation loop around the thickener to prevent settling of the solids and plugging of the lines. The feed to the filter is taken as a bleed from this recirculation line.

The filter operates on overflow using an adjustable weir to set the proper slurry level in the filter tub. The overflow slurry is returned to the thickener. The solids cake is washed on the filter using a series of wash water sections. Solids from the filter cake are discharged to a weigh belt conveyor for transport to a waste processing system. The mixed filtrate and wash liquor from the filter are returned to the thickener.

d. Raw Materials Preparation

Two raw materials are required for operation of the system--limestone for absorbent regeneration and soda ash to make up for losses of sodium salts in the waste filter cake.

Limestone will be received, stored, and, under normal conditions, fed to the system in a dry form. The flexibility exists for feeding it either to the first or second reactor tank. For operation with limestone slurry, limestone would be fed to the first reactor tank along with water, and the scrubber bleed diverted to the second reactor tank.

Dense soda ash will be used to make up for sodium salt losses. Provisions have been made to prepare soda ash solutions using either river water or clarified liquor from the thickener hold tank. The flexibility also exists to feed soda ash solution to the thickener, thickener hold tank, absorber recycle, or venturi scrubber recycle.

3. Waste Processing System

The waste processing system will be installed and operated as part of a separate test program funded by EPRI. It is briefly discussed in this report to indicate the overall plant operation.

The moist waste cake discharged from the vacuum filter will be mixed in a pug mill with fly ash and lime. The amount of ash added will range between 0.5 to 1.5 parts of ash per part of dry solids in the cake. Lime will be added at a rate of 2 to 5 wt.% of the dry solids contained in the mix of cake and fly ash.

The mix discharged by the pug mill will be loaded onto trucks and transported to a sectioned landfill area. Three test areas and one off-spec fill area will be used for disposing of the processed limestone dual alkali waste. Each of the test areas will be filled with a different waste material mix involving different combinations of filter cake, fly ash, and lime. Any waste material not meeting specifications will be disposed of in the off-spec area.

C. SYSTEM DESIGN

1. Design Basis

The modified limestone dual alkali process has been designed to operate as a closed-loop system in accordance with the following criteria:

- Operation in a "concentrated" alkali mode (>1.0 M active alkali);
- Operation with minimum sodium consumption;
- Operation with minimum calcium consumption (95% limestone utilization);
- Removal of more than 95% of the incoming sulfur dioxide;
- Production of a solid waste filter cake containing 55 to 65% insoluble solids (calcium sulfite/sulfate); and
- Reliable system operation with normal variation of:
  - flue gas composition from the power plant,
  - boiler load,
  - electrostatic precipitator operation, and
  - limestone quality.

During the test program, the boilers at Scholz will be fired with coal containing 2.9 to 3.4 wt.% sulfur. The corresponding SO_2 levels in the boiler flue gas have been estimated to be on the order of 1900 to 2300 ppm. During the test program, it may be desirable to increase these

levels by injecting SO_2 into the flue gas stream before it enters the scrubber. The modification of the dual alkali system has been designed to accommodate operation under normal conditions without SO_2 injection, as well as operation with SO_2 injection. The design bases for each of these operating modes are given in Table III-1. The corresponding material balances are discussed below.

2. Material Balances

The material balances for the dual alkali system were developed to meet the specific design bases given in Table III-1 as well as the system design criteria referred to above. These material balances represent flow rates and compositions averaged over a 24-hour period of operation. The basic difference between the actual operation of the system and the 24-hour average is the operation of the filter. The waste generated in 24 hours can be optimally removed from the system by operating the filter 7 to 12 hours a day. Thus, the flow rates of the streams associated with the filtering operation change during a daily operation depending on whether the filter is on or off line.

In addition, the material balances do not account for non-process related system inputs and outputs; namely, open tank evaporation, rainfall, and pump seal water. In order to determine the required surge capacity and actual makeup water requirements, these factors have been incorporated in the volume balance discussed in a subsequent section of this chapter.

a. Normal Operation

The basis for the material balance under normal operation is presented in Table III-2. The material balance, providing component and total flow rates for each process stream, is given in Table III-3. Although independently identified in this table, each process stream may be correlated to the process flow diagram in CEA drawing 8101-1-1 by the stream numbers.

The flue gas taken by the dual alkali system (40,000 scfm) is equivalent to a load of approximately 18 Mw, with an SO_2 level of 2200 ppm (dry basis). Removal of 95% of the SO_2 present in the inlet gas generates about 53 lb/min of moist waste cake. The solids content of this filter cake for the purposes of this material balance is assumed to be about 55%. Limestone would be added to the system at a rate of 23 lb/min, (0.986 moles of $CaCO_3$/mole of ΔSO_2) and soda ash is added at a rate of less than 1 lb/min (0.035 moles Na_2CO_3/mole of ΔSO_2).

Internally, the regenerated solution is fed to the absorber/scrubber module at a total rate of about 167 gpm (corresponding to an L/G = 3.1). The active sodium concentration in this solution is 1.7 M, with a $[HSO_3^-]$ to $[SO_3^=]$ ratio of 1.34. The spent solution with a $[HSO_3^-]$ to $[SO_3^=]$ ratio of 3.0 is bled to the regeneration reactors at about 145 gpm. The remaining 22 gpm are evaporated in bringing the incoming hot flue gas to saturation.

TABLE III-1

LIMESTONE DUAL ALKALI SYSTEM
DESIGN BASIS

	Normal Operation	Additional SO_2 Injection
Inlet Gas Flow	40,000 scfm (dry)	45,000 scfm (dry)
Inlet SO_2	2,200 ppm (dry)	2,650 ppm (dry)
Inlet Chloride (Cl^-)	0.12 lb/min.	0.12 lb/min.
Inlet O_2	5 vol.%	5 vol.%
SO_2 Removal Efficiency	95%	95%

TABLE III-2

BASIS FOR PROCESS MATERIAL BALANCES
NORMAL OPERATION

Inlet Gas:	
Flow Rate	40,000 scfm (dry)
SO_2	14.9 lb/min. (2,200 ppm dry)
Cl^-	0.12 lb/min.
O_2	5 vol.%
Moisture	6.0 wt.%
Removal Efficiency:	
SO_2	95%
Cl^-	100%
Absorber Feed Concentration:	
Na^+ associated with	
$SO_3^=$	1.02 M
HSO_3^-	0.68 M
Oxidation and Evaporation:	
Oxidation	7 mole % of ΔSO_2
Evaporation in Absorber	22 gpm
Soda Ash and Limestone Feed:	
Soda Ash Feed Rate	0.07 mole Na^+/mole ΔSO_2
Limestone Purity	95 wt.%
Limestone Feed Rate	0.986 mole available $CaCO_3$/mole ΔSO_2
Waste Solids:	
Wash Ratio	4 displacement washes
Insoluble Solids	55 wt.%

TABLE III-3

MATERIAL BALANCE - NORMAL OPERATION

Stream No.	Fan Inlet 1	Venturi Inlet 2	Absorber Outlet 3	Stack Outlet 4	Combustion Air 5
Bone Dry Gas, lb/min.	3151.4	3151.4	3151.4	3207.0	
SO_2, lb/min.	14.9	14.9	0.75	0.79	
H_2O, lb/min.	202.1	202.1	385.6	397.6	
Total, lb/min.	3368.4	3368.4	3537.8	3605.4	
Temperature, °F	360	370	133	173	70
Pressure, inches WG	-1	+19	+1	+1	+20
Volume, acfm	63,200	61,000	54,500	59,400	715
SO_2, ppm dry	2,200	2,200	111	115	
Moisture, wt.%	6.0	6.0	10.9	11.0	

TABLE III-3
(continued)

MATERIAL BALANCE - NORMAL OPERATION

Stream No.	Feed Forward 6	Limestone Feed 8	Reactor Effluent 9	Thickener Underflow 10	Thickener Overflow 11	Filtrate Recycle 12	Filter Cake 13
H_2O, lb/min.	1307.6		1121.9	65.4	1195.0	138.5	22.9
Na_2CO_3, lb/min.							
Na_2SO_3, lb/min.	88.5		86.8	5.1	86.6	4.9	0.2
$NaHSO_3$, lb/min.	97.9		99.8	5.8	99.5	5.5	0.3
Na_2SO_4, lb/min.	127.9		128.3	7.5	127.9	7.1	0.4
NaCl, lb/min.	71.3		71.5	4.2	71.3	4.0	0.2
$CaSO_3 \cdot 1/2\ H_2O$, lb/min.			25.9	25.9			25.9
$CaSO_4 \cdot 1/2\ H_2O$, lb/min.			1.9	1.9			1.9
$CaCO_3$, lb/min.		21.8	0.4	0.4			0.4
Inerts, lb/min.		1.1	1.1	1.1			1.1
Total, lb/min.	1693.2	22.9	1537.6	117.3	1580.3	160.0	53.3
Insolubles, wt.%	0.0	100.0	1.9	25.0	0.0	0.0	55.0
Flow, gpm	167	---	146	10	153	17	---

TABLE III-3
(continued)

MATERIAL BALANCE - NORMAL OPERATION

	Soda Ash Solution	Makeup Water	Soda Ash Feed	Soda Ash Water	Scrubber Bleed	Limestone Water	Cake Wash Water
Stream No.	14	15	16	17	35	39	40
H_2O, lb/min.	3.3	109.2		3.3	1120.0	0.0	96.0
Na_2CO_3, lb/min.	0.8		0.8				
Na_2SO_3, lb/min.					58.3		
$NaHSO_3$, lb/min.					144.2		
Na_2SO_4, lb/min.					130.1		
NaCl, lb/min.					71.5		
Total, lb/min.	4.1	109.2	0.8	3.3	1524.1	0.0	96.0
Insolubles, wt.%	0.0	0.0	100.0	0.0	0.0	0.0	0.0
Flow, gpm	0.4	13.0	---	0.4	145	0	11.5

The reactor effluent is a slurry containing 2 wt.% insoluble solids. The solids concentration is increased to at least 25 wt.% in the thickener and finally to a minimum of 55 wt.% in the filter cake. In order to reduce the loss of soluble sodium salts contained in the cake liquor, the filter cake is washed with 12 gpm of water (equivalent to displacing the liquor in the final cake four times). The filtrate, along with the wash water, is returned to the thickener.

In this material balance, the soda ash solution is assumed to be fed to the thickener hold tank, although the capability exists for feeding it to the thickener, absorber, or venturi scrubber. Similarly, river water is assumed to be used in the preparation of the soda ash solution but the capability exists for using clarified process liquor from the thickener hold tank. The rate of soda ash makeup required to replace the sodium value lost in the cake is 0.8 lb/min, equivalent to 0.035 moles of Na_2CO_3 per mole of SO_2 removed.

Based on the total soda ash and limestone feed rates, the overall system alkali stoichiometry is estimated to be 1.014 (moles of available $CaCO_3$ + Na_2CO_3)/(mole of ΔSO_2 + ΔCl_2).

b. Maximum Operation with SO_2 Injection

The assumptions made in developing the material balance for the operation of the system at maximum gas flow with an additional injection of SO_2 are listed in Table III-4. The detailed material balance is presented in Table III-5.

At maximum load, the flue gas treated by the system amounts to 45,000 scfm, roughly equivalent to a 20 Mw boiler load. The SO_2 loading has been increased by about 20% to 2650 ppm by the injection of SO_2. Removal of 95% of the incoming SO_2 produces about 77 lb/min of moist cake (at 55 wt.% dry solids). Limestone is added to the system at a rate of 34 lb/min (1.010 moles $CaCO_3$/mole of ΔSO_2), and soda ash is added at a rate of 1.3 lb/min (0.040 moles Na_2CO_3/mole ΔSO_2).

The regenerated scrubbing solution is fed forward to the absorber/scrubber module at a rate of 205 gpm (equivalent to an L/G = 4.5). The liquor feed to the absorber has an active sodium concentration of 1.7 M and an $[HSO_3^-]$ to $[SO_3^=]$ ratio of 1.34. The spent solution, with an $[HSO_3^-]$ to $[SO_3^=]$ ratio of 3.2, is bled to the regeneration reactors at a rate of 180 gpm. The difference in flows, about 25 gpm, represents the water evaporated to saturate the incoming flue gas.

The solids concentration in the resulting slurries are consistent with those generated in normal operation. The reactor effluent contains 2.2 wt.% solids, the thickener underflow at least 25% and the filter cake a minimum of 55%. The cake is washed with 16.5 gpm equivalent to a 4 displacement wash ratio.

TABLE III-4

BASIS FOR PROCESS MATERIAL BALANCES
SO_2 INJECTION

Inlet Gas:	
Flowrate	45,000 scfm (dry)
SO_2	20.2 lbs/min (2,650 ppm dry)
Cl^-	0.12 lbs/min
O_2	5 vol.%
Moisture	6.0 wt.%
Removal Efficiency:	
SO_2	95%
Cl^-	100%
Absorber Feed Concentration:	
Na^+ associated with	
$SO_3^=$	1.02 M
HSO_3^-	0.68 M
Oxidation and Evaporation:	
Oxidation	10 mole % of ΔSO_2
Evaporation in Absorber	24.7 gpm
Soda Ash and Limestone Feed:	
Soda ash feed rate	0.081 moles Na^+/mole ΔSO_2
Limestone purity	90%
Limestone feed rate	1.010 moles available $CaCO_3$/mole ΔSO_2
Waste Solids:	
Wash ratio	4 displacement washes
Insoluble solids	55 wt.%

TABLE III-5

MATERIAL BALANCE - SO_2 INJECTION

Stream No.	Fan Inlet 1	Venturi Inlet 2	Absorber Outlet 3	Stack Outlet 4	Combustion Air 5
Bone Dry Gas, lb/min.	3540.7	3540.7	3540.7	3603.2	
SO_2, lb/min.	20.2	20.2	1.0	1.1	
H_2O, lb/min.	227.4	227.4	433.4	446.1	
Total, lb/min.	3788.3	3788.3	3975.1	4050.4	
Temperature, °F	360	370	133	173	70
Pressure, inches WG	-1	+19	+1	+1	+20
Volume, acfm	78,700	75,900	61,300	66,700	800
SO_2, ppm dry	2,660	2,660	130	140	
Moisture, wt.%	6.0	6.0	10.9	11.0	

TABLE III-5
(continued)

MATERIAL BALANCE - SO_2 INJECTION

Stream No.	Feed Forward 6	Limestone Feed 8	Reactor Effluent 9	Thickener Underflow 10	Thickener Overflow 11	Filtrate Recycle 12	Filter Cake 13
H_2O, lbs/min	1594.0		1385.1	92.0	1490.6	197.6	32.9
Na_2CO_3, lbs/min							
Na_2SO_3, lbs/min	109.7		107.0	7.1	106.7	6.8	0.3
$NaHSO_3$, lbs/min	121.9		124.8	8.3	124.4	7.9	0.4
Na_2SO_4, lbs/min	229.2		230.0	15.3	229.2	14.5	0.8
NaCl, lbs/min	63.0		63.2	4.2	63.0	4.0	0.2
$CaSO_3 \cdot 1/2\ H_2O$, lbs/min			33.9	33.9			33.9
$CaSO_4 \cdot 1/2\ H_2O$, lbs/min			3.6	3.6			3.6
$CaCO_3$, lbs/min		30.2	1.5	1.5			1.5
Inerts, lbs/min		3.4	3.4	3.4			3.4
Total, lbs/min	2117.8	33.6	1952.5	169.3	2013.9	230.8	77.0
Insolubles, wt.%	0.0	100.0	2.2	25.0	0.0	0.0	55.0
Flow, gpm	205	--	181	14	192	25	--

TABLE III-5
(continued)

MATERIAL BALANCE - SO_2 INJECTION

Stream No.	Soda Ash Solution 14	Make-up Water 15	Soda Ash Feed 16	Soda Ash Water 17	Scrubber Bleed 35	Limestone Water 39	Cake Wash Water 40
H_2O, lbs/min	5.1	98.1		5.1	1382.5	0.0	138.4
Na_2CO_3, lbs/min	1.3		1.3				
Na_2SO_3, lbs/min					67.8		
$NaHSO_3$, lbs/min					184.5		
Na_2SO_4, lbs/min					233.5		
NaCl, lbs/min					63.2		
Total, lbs/min	6.4	98.1	1.3	5.1	1931.5	0.0	138.4
Insolubles, lbs/min	0.0	0.0	100.0	0.0	0.0	0.0	0.0
Flow, gpm	0.6	11.8	--	0.6	180	0.0	16.6

The soda ash makeup rate is 1.3 lb/min, corresponding to 0.04 mole of Na_2CO_3 per mole of SO_2 removed. Based on the total soda ash and limestone feed rates, the overall alkali stoichiometry for the system is estimated at 1.044 (moles of available $CaCO_3$ + Na_2CO_3)/(mole of ΔSO_2 + ΔCl_2).

3. Volume Balances

Ordinarily, the water lost from the dual alkali system due to evaporation and cake losses exceeds the required process water inputs, and a water balance (as well as system volume balance) is maintained by adding water to the thickener hold tank on liquid level. If, however, the filter is operated for only a fraction of a day (7 to 12 hours), the actual rate of cake generation is substantially higher than the 24-hour average given in the material balance. Similarly, the cake wash rate increases correspondingly, reducing the makeup water requirement to maintain a constant level in the thickener hold tank. Under extreme conditions of high filtration rates and excessive rainfall, even a total curtailment in the makeup water to the hold tank may result in a temporary increase in the overall system volume. Provisions are therefore made in the process design for surge capacity based upon expected worst-case system volume imbalances, taking into account all process and non-process (open tank evaporation, rain, etc.) inputs and outputs to the system. Water and volume balances have been calculated to correspond with each of the process material balances--normal operation and maximum operation with SO_2 injections. The assumptions made in developing the water and system volume balances are given in Table III-6.

a. Normal Operation

During normal operation, the dual alkali system generates insoluble waste solids at a rate of 29.3 lb/min, or 21 tons/day, which is equivalent to 38 tons/day of moist cake (55 wt.% solids). It may be desirable though to operate the filter only one to one and one-half shifts per day to allow for disposal operations during daylight hours. Assuming operation of the filter for 7 hours per day, filter cake would be produced at a rate of about 3 tons of dry solids per hour, or 5.4 tons/hour of moist cake (55 wt.% solids).

The short period of filter operation dictates that the daily average cake wash water requirement of 11.5 gpm (see material balance, Table III-3, and water and volume balance, Table III-7) must be supplied in only 7 hours or at a rate of 39.5 gpm. This large inflow of water causes a temporary increase in the system volume. The surge capacity required in the system to contain the excess volume during operation of the filter amounts to 1300 gallons on a normal day and 5500 gallons on a rainy day. The liquor stored in the surge tank is later used instead of fresh makeup water to the thickener hold tank.

TABLE III-6

ASSUMPTIONS FOR THE WATER AND SYSTEM VOLUME BALANCES
(Expressed in gpm)

	Normal Operation		SO_2 Injection	
	24 Hour Average	Filter on (7 hr/day)	24 Hour Average	Filter On (11 hr/day)
Pump Seals (Input)	6	6	6	6
Process Evaporation (Absorbers)	22	22	24.5	24.5
Open Tank Evaporation				
• Normal days				
Net evaporation[a]	7	7	7	7
• Rainy days				
Net evaporation	4[b]	—	4[b]	—
Net rain input	—	3[c]	—	1.5[d]

[a]No rainfall

[b]Rainfall = 3 in/day

[c]Rainfall = 3 in/7 hours

[d]Rainfall = 4 in/11 hours

TABLE III-7

SYSTEM VOLUME BALANCE AND SURGE CAPACITY REQUIREMENTS
NORMAL OPERATION

	24 h Average Normal Day	24 h Average Rainy Day	Filter On (7 h) Normal Day	Filter On (7 h) Rainy Day
Input				
Limestone	1	1	1	1
Soda Ash Solution	0.5	0.5	0.5	0.5
Make-up Water	14	11		
Cake Wash Water	11.5	11.5	39.5	39.5
Pump Seals	6	6	6	6
Rain (net of evaporation)				3
Total	33 gpm	30 gpm	47 gpm	50 gpm
Output				
Evaporation-Absorbers	22	22	22	22
Evaporation-Open Tanks (net of rain)	7	4	7	
Waste Cake	4	4	15	15
Total	33 gpm	30 gpm	44 gpm	37 gpm
Surge Capacity Required	---	---	1300 gals	5500 gals

b. SO_2 Injection

The water and volume balance for the SO_2 injection case are developed in a manner analogous to that for the normal operation. The results are shown in Table III-8.

The system generates insoluble waste solids at a rate of 42.4 lb/min, or 30.5 tons/day, which is equivalent to 55.5 tons/day of moist cake (55 wt.% solids). The filter requires approximately 11 hours to remove these solids. During the filter operation, 36 gpm of water are used to wash the cake. The corresponding surge capacity required is 5300 gallons on a rainy day. No surge capacity is needed on normal dry days.

TABLE III-8

SYSTEM VOLUME BALANCE AND SURGE CAPACITY REQUIREMENTS
SO_2 INJECTION

	24 h Average Normal Day	24 h Average Rainy Day	Filter On (11 h) Normal Day	Filter On (11 h) Rainy Day
Input				
Limestone	1.5	1.5	1.5	1.5
Soda Ash Solution	1	1	1	1
Make-up Water	13	10	0.5	
Cake Wash Water	16.5	16.5	36	36
Pump Seals	6	6	6	6
Rain (net of evaporation)				1.5
Total	38 gpm	35 gpm	45 gpm	46 gpm
Output				
Evaporation-Absorbers	24.5	24.5	24.5	24.5
Evaporation-Open Tanks (net of rain)	7	4	7	
Waste Cake	6.5	6.5	13.5	13.5
Total	38 gpm	35 gpm	45 gpm	38 gpm
Surge Capacity Required	---	---	---	5300 gals

IV. TEST PROGRAM[1]

The startup and breakin testing of the system will last approximately two months; immediately following, the system will undergo a six month period of testing to characterize its operations. This test program is aimed at evaluating the following aspects of the limestone dual alkali system performance:

- SO_2 removal capabilities;
- Consumption of raw materials--soda ash, limestone, and water--not only as a measure of the efficiency of the system in utilizing these materials in the removal of SO_2 but also as a measure of the ability of the system to maintain a closed-loop operation;
- Consumption of energy--electrical power, and fuel for the gas reheater;
- Levels of sulfite oxidation experienced by the system as well as the ability of the system to tolerate such levels of oxidation;
- Waste cake properties--dewatering, handling, and washability;
- Particulate matter removal (during periods of planned precipitator turndown) by the system and the effect of ash on process chemistry; and
- Sulfate emissions in the processed flue gas.

The test program will be conducted in two phases. During the first phase, a baseline for the operation of the system will be established. The second phase will consist of variational tests of the system.

A. SCHEDULE

It is anticipated that the flue gas desulfurization (FGD) section of the system will be ready for startup in March 1980; the waste disposal system, however, will not be completed until the end of June 1980. Given the need to coordinate the operating phases of both the FGD and the waste disposal test programs, the start of the test programs has been delayed until early August 1980. The startup and breakin testing for the FGD system will take

[1]This test program represents the original test program issued in draft form in March 1980. For reasons indicated in Sections I and II, the testing period of this program has been reduced from six to two months. The system was started up in August 1980 and formal testing began in February 1981.

place in the months of June and July 1980. Phase I, the baseline testing, will last from early August to the end of September 1980; and Phase II, the variational testing, will last from early October 1980 to the end of January 1981.

B. TEST PROGRAM PHASES[2]

1. Startup and Breakin Testing

The startup and breakin testing of the system will be performed during the months of June and July 1980. During this period, not only will the system be started up, but it will also undergo preliminary testing. The information obtained in these tests will be used to optimize and finalize the control set points, and to verify the suitability of the testing and analytical procedures. Furthermore, the results of the tests will be used to make any necessary adjustments to the system, to the system operation, and to the test plan.

The preliminary tests will include the determination of: (a) SO_2 removal efficiency at various gas loadings, inlet SO_2 concentrations, and scrubber operating pH's; (b) regeneration efficiency as a function of reactor configuration and type of limestone--if possible the effect of limestone particle size will also be evaluated; and (c) waste solids properties and conditions for operation of the thickener, filter, and cake wash system. A list of the projected operating conditions during the startup and the test program is given in Table IV-1.

2. Phase I - Baseline Testing

In previous pilot plant and laboratory studies, the general effect of process parameters and operating conditions on the performance of limestone-based dual alkali systems was determined. The process parameters/operating conditions and the performance variables that they affect are listed in Table IV-2. The baseline testing for this program will focus on confirmation and additional characterization of these relationships. Most important, a baseline testing will provide a quantitative evaluation of the overall system performance for an extended period for at least one (nominal) set of process parameters/operating conditions. Subsequent variations in selected process parameters/operating conditions and their effect on the performance of the system will be studied in the second phase of this program.

3. Phase II - Variational Testing

This phase of the test program will last approximately four months. The results of this variational test will provide the technical information needed to determine the range of application and limitations of the limestone dual alkali technology. The various conditions under which the system will be operated will involve, primarily, variations in those conditions which could be expected to be encountered in full-scale applications:

[2]See footnote on previous page

TABLE IV-1

PROTOTYPE TEST PROGRAM

PROJECTED OPERATING CONDITIONS

Period	Nominal Load (Mw)	Nominal Inlet SO_2 (ppm dry)	Limestone Type*	Reactor Configur.	Inlet O_2 Vol.%	Particulate Loading	Comments
Startup							
2 months	10-20	1800-2800	A/B	Variable	4.5-6	None	Finalize control set points. Evaluate testing and analytical process. Perform preliminary evaluation of the system response to boiler load, SO_2 load, reactor configuration, and type of limestone. Characterize filter performance. Check gas flow vs. fan amperage correlations.
Phase I							
2 months	18-20	2000-2500	A or B	Standard	4.5-6	None	Baseline Testing
Phase II							
4 months							
3 wks	10-20	2000-2500	A or B	Standard	4.5-6	None	Boiler (Scrubber) Load Testing
1-2 wks	18	1500-2000	A or B	Standard	4.5-6	None	SO_2 Load Testing
1-2 wks	18	2000-2800	A or B	Standard	4.5-6	None	SO_2 Load Testing
2 wks	18	2000-2500	A or B	Variable	4.5-6	None	Reactor Testing
2 wks	18	2000-2500	A or B	Standard	4.5-10	None	Scrubber Oxidation
2-3 wks	18	2000-2500	A1 or B1	Standard	4.5-6	None	Limestone Testing
2-3 wks	18	2000-2500	C	Standard	4.5-6	None	Limestone Testing
2 wks	18	2000-2500	A,B, or C	Standard	4.5-6	Variable	Particulate Load Testing

*A,B,C--Different types of limestone. A1,B1--Different particle size.

TABLE IV-2

PROCESS PARAMETERS AND OTHER CONDITIONS WHICH AFFECT THE PERFORMANCE OF THE LIMESTONE-BASED DUAL ALKALI SYSTEM

Performance Variable	Parameters/Conditions
SO_2 Removal Efficiency	Inlet SO_2 concentration/gas flow Scrubber configuration (venturi + absorber, venturi alone) Scrubbing liquor pH (inlet and outlet) Venturi pressure drop Active sodium concentration
Limestone Consumption	Inlet SO_2 and O_2 concentration/gas flow Type of limestone and particle size Reactor configuration Reactor hold-up time
Soda Ash Consumption	Inlet SO_2 and O_2 concentration/gas flow Liquor entrainment in scrubber section Cake wash efficiency/wash ratio
Energy Consumption	Gas flow Inlet SO_2 concentration
Oxidation	Inlet O_2 concentration Scrubber configuration Scrubber temperature
Cake Properties	Type of limestone Reactor configuration Reactor hold-up time Reactor pH Filter operation
Particulate Removal	Inlet particulate concentration/gas flow Scrubber configuration
Process Operability	Operational upsets in system conditions (equipment failure, operator error, etc) Scaling, pluggage, spills

TABLE IV-3

SYSTEM PERFORMANCE EVALUATION - DATA REQUIREMENTS

Data Requirements	Directly Measured	Derived From
• Volume of gas treated		Fan amperage & calibration curves
• Amount of SO_2 removed		Volume of gas treated, inlet & outlet SO_2 concentrations
• Quantity of cake generated	✓	
• Cake quality		Moisture content, insoluble and soluble salts composition
• Cake wash efficiency		Filter operating conditions, wash water composition, soluble salts content in cake and in underflow slurry liquor
• Limestone consumption	✓	Check: cake generated & insoluble salts content; inventory & deliveries
• Limestone utilization		Cake composition--insoluble salts Limestone consumption and SO_2 removed
• Soda ash consumption	✓	Check: cake generated & soluble salts content; inventory & deliveries
• Level of oxidation		Solution composition and flows: feed forward to absorber/ scrubber, scrubber bleed, reactor effluent Filter cake compositions
• Reactor solids quality		Solids content in slurry Settling characteristics Filter cake solids content
• Scrubber pH--inlet & outlet	✓	
• Venturi and absorber pressure drop	✓	
• Reactor effluent pH	✓	
• Reactor configuration & hold-up time		Reactor train volume, reactor effluent flow
• Utilities requirement: water, electricity, reheater fuel	✓	
Additional Data for Specific Periods		
• Amount of particulate matter removed		Volume of gas treated; inlet & outlet particle concentrations
• Sulfate emissions in treated gas	✓	

- boiler load fluctuations,
- inlet SO_2 concentrations,
- type of limestone used and/or particle size,
- reactor configuration and residence time,
- boiler excess air (O_2 concentration in inlet flue gas), and
- particulate loading.

The operation of the system, while evaluating the effect of each of the above variables, will constitute a discrete testing period. Each testing period will last approximately two to three weeks unless the operating conditions or other situations dictate a more reasonable or convenient period. A summary of the various projected operating conditions is given in the outline of the overall test program presented earlier in Table IV-1.

C. DATA REQUIREMENTS

The evaluation of the limestone dual alkali system during any one operating period requires material and energy balances as well as information on selected physical and chemical operating conditions. These data requirements are listed in Table IV-3. Part of the data is amenable to direct measurements and part of it must be derived indirectly from other data. For this latter case, the additional data required are also listed in Table IV-3.

In general, a summary of the system performance will be developed for each test period. The data, however, will be analyzed on a daily basis; and a significant fraction of it will be collected every two hours. Thus, sufficiently detailed information will be available for the development of specific performance correlations (e.g., SO_2 removal vs. scrubber pH).

The frequency procedures, instrumentation, and analyses needed for gathering the experimental data are described in the next section. The reduction of data is also discussed in a subsequent section.

D. DATA GATHERING

Collection of the test data for evaluating the system will basically involve monitoring of process instrumentation and chemical analyses of samples from process vessels and streams.

1. Process Instrumentation

The system has been supplied with sufficient instrumentation to allow the monitoring of the important variables in the operation of the system. Temperature sensors are provided for the fan bearings; scrubber, absorber, and reheater inlet and outlet gas streams; scrubber and absorber liquor recycle; thickener overflow and soda ash solution. Pressure drops are monitored across the scrubber and the absorber. SO_2 concentrations in the gas streams are measured before entering the dual alkali system and after exiting the reheater. PH probes are supplied for the absorber and

scrubber liquor recycle streams, and for the last two reactors in the 5-tank reactor system. With the exception of the thickener overflow and underflow, all other major liquid/slurry process streams are equipped with flow meters. Liquid level is monitored in the scrubber and absorber tanks; the last tank in the reactor system; the thickener hold tank; the soda ash solution tank; and the surge tank. Weigh feeders control the addition of limestone and soda ash and monitor the generation of waste cake. The water consumption by the system is monitored by flow totalizers. An ammeter is provided measuring the fan amperage.

The output of most of these instruments is displayed on the control room panel, either in a chart recorder, analog indicator, or in a digital display from a data logger; thus facilitating the data collection duties of the operators. All data displayed on the control room panel will be recorded every two hours; some field data will also be recorded every two hours and some field data will be recorded daily or on special occasions. The recommended data log format is presented in Table IV-4.

2. Process Sampling and Sample Analyses

Samples will be taken from all major process streams for chemical analyses. The sampling locations are shown in Figure IV-1. The sampling frequency is summarized in Table IV-5. The frequency may be adjusted to accommodate special testing if deemed necessary to further characterize a given process unit or system chemistry.

Slurry samples will require separation of solids and liquor prior to chemical analyses. A list of the required phase separations and analyses for the characterization of each process sample is given in Table IV-6. A number of chemical analyses indicated in this table may not be required under most operating conditions. These have been shown in parentheses. For example, there may be very little detectable iron in the process samples. If no evidence of this species is found in the initial samples or if the concentrations are very low, then these analyses will be performed only intermittently on a few selected samples as required.

Not all samples will be fully analyzed; rather, only the critical analyses and separations will be performed with the remainder of the sample inerted and "stockpiled." A complete analysis can then be performed on an as-needed basis.

The number of analyses which will ultimately be performed will be determined by the needs of the test program as it progresses and the limitations of finite analytical facilities and manpower. A complete set of analyses on all samples would be burdensome, expensive, and to a large extent, pointless, since it would not be necessary to close material balances or perform complete characterizations on a daily basis around each process section.

TABLE IV-4

CEA/ADL LIMESTONE DUAL ALKALI SYSTEM

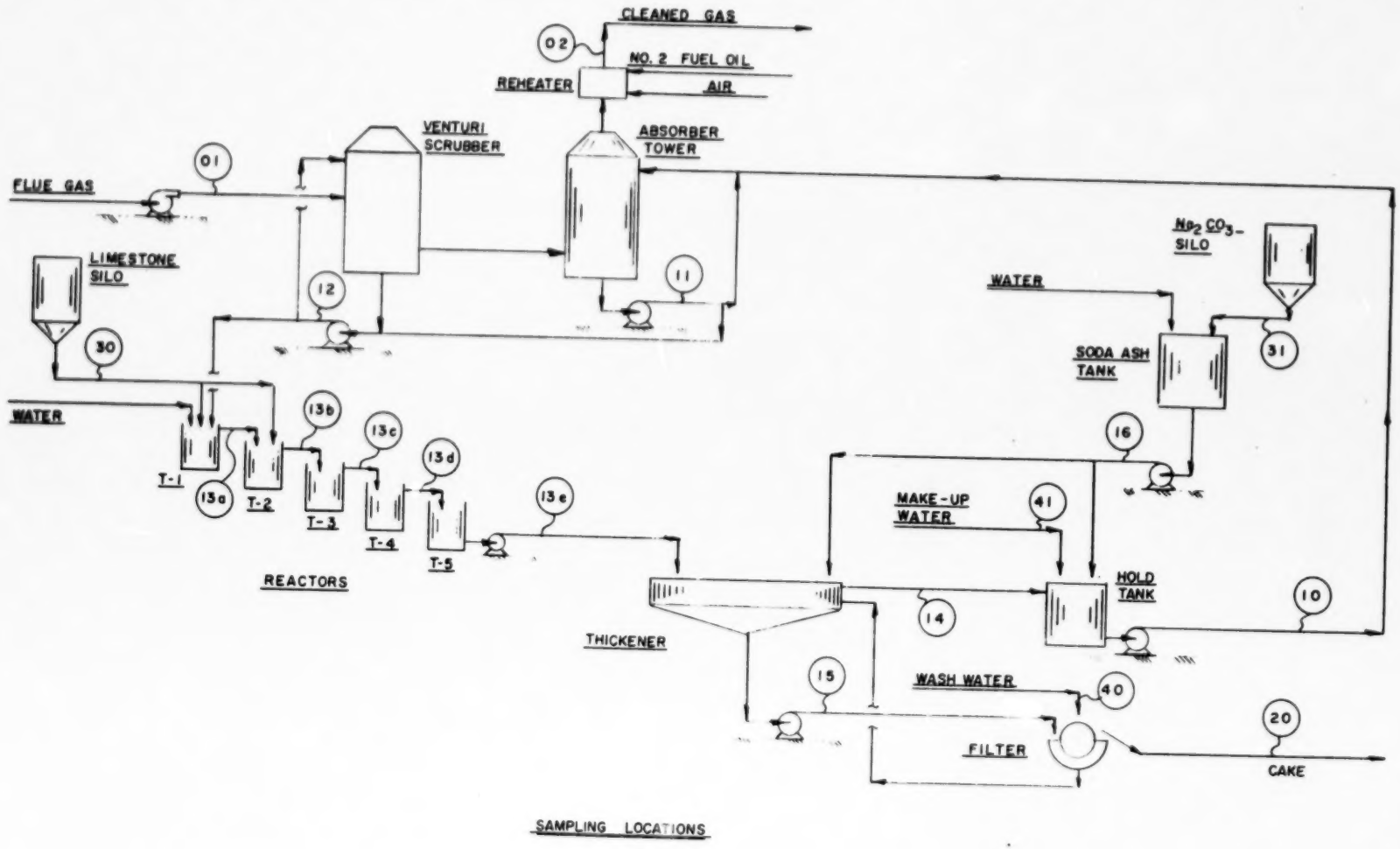
SCHOLZ PLANT - GULF POWER CO.

Date _____

Time		0100	0300	0500	0700	0900	1100	1300	1500	1700	1900	2100	2300	
Operators Initials														
	Tag													
Temperatures														
Fan Inboard	T-1													
Scrub. Inlet Gas	T-2													
Scrub. Outlet Gas	T-3													
Abs. Outlet Gas	T-4													
Reheated Gas	T-5													
Scrub. Recycle	T-6													
Abs. Recycle	T-8													
Soda Ash Solution	T-9													
Fan Outboard	T-20													
Analyzers														
Inlet Gas SO_2	A-1													
Outlet Gas SO_2	A-2													
Scrub. Recycle pH	A-3													
Abs. Recycle pH	A-4													
Reactor T-4	A-5													
Reactor T-5	A-6/A-7													
Pressures														
Scrub. Drop	Δ-1													
Abs. Drop	Δ-2													
Fan Inlet	P-3													
Scrub. Inlet	P-4													
Abs. Outlet	P-5													
Liquid Flow														
Scrub. Bleed	F-1													
Scrub. Recycle	F-3													
Abs. Recycle	F-5													
Regen. Feed-Forward	F-7													
Soda Ash Solution	F-14													
Reactor Effluent	F-16													
Make-up Water-Soda Ash	F-19													
Gas Flow														
Fan Amps	X-8													
Damper Position	H-1													
SO_2 Injection														
Tank Levels														
Scrubber	L-1													
Absorber	L-2													
Thickener Hold Tank	L-4													
Reactor T-5	L-5													
Soda Ash Solution	L-6													
Surge Tank	L-7													
Dry Feeders														
Dry Limestone	W-18													
Dry Soda Ash	W-19													
Field Data														
Abs. Flow to Scrub.	F-10													
Water-Filter Sprays														
Water-Vacuum Pump														
Water-Thick.Hold Tank	F-13													
Thickener Rake Height														

DAILY READINGS

	Reading	Time		Reading	Time
River Water Totalizer (Gals)			Dry Limestone Totalizer (lbs)		
Well Water Totalizer (Gals)			Power (kw-hr)		
Filter Wash Totalizer (Gals)			Fuel Oil (Gals)		
Hold Tank Make-up Totalizer (Gals)			Filter Cake Conveyor (lbs)		
Dry Soda Ash Totalizer (lbs)					



SAMPLING LOCATIONS

FIGURE IV-1

TABLE IV-5

SAMPLING REQUIREMENTS

Sample Point	Designation	Sample Location	Sampling Frequency
Gas Streams			
01	Inlet Flue Gas (SO_2)	Booster Fan Discharge	Monitored continuously
02	Cleaned Gas (SO_2)	Reheater Exhaust	by process instrumentation
Process Liquor/Slurry			
10	Tray Tower Regenerated Feed (Feed Forward)	Pump Discharge	D
11	Tray Tower Bleed	Pump Discharge	D
12	Venturi Scrubber Bleed	Pump Discharge	D
13a	Reactor Effluent	Tank 1	Samples from intermediate reactors will be taken as necessary
13b		Tank 2	
13c		Tank 3	
13d		Tank 4	
13e		Tank 5	D
14	Thickener Overflow	Overflow Weir	D
15	Thickener Underflow Slurry	Pump Discharge	D
16	Soda Ash Solution	Tank or Pump Discharge	W
System Discharge			
20	Filter Cake	Dock Blade	D
Raw Materials			
30	Limestone Feed (Dry)	Weigh Belt	D
31	Soda Ash (Dry)	Weigh Belt	D
Water Supply			
40	Make Up Well Water		M
41	Make Up River Water		M

TABLE IV-6

ANALYSES OF PROCESS SAMPLES*

Sample Point	Designation	Sample Location	Physical Test			Analyses of Solutions											Analyses of Solids						
			Specific Gravity	Solids Settling	Liquid/Solid Separation	TOS	$SO_4^=$	H^+	Total CO_2	Cl^-	NO_3^-	Na^+	Ca	Mg	pH	Fe	Other	TOS	$SO_4^=$	$CO_3^=$	Cl^-	Ca	Mg	Na^+
10	Tray Tower Regenerated Feed (Feed Forward)	Thickener Hold Tank Pump Discharge	X			X	X	X	X	(X)	(X)	(X)	X	X	X	(X)								
11	Tray Tower Bleed	Pump Discharge				X	X		X	(X)	(X)	(X)	X	X	X	(X)								
12	Venturi Scrubber Bleed	Pump Discharge	X			(X)	(X)		X	(X)	(X)	(X)	X	X	X	(X)								
13a	Reactors Effluent	Tank		(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)		(X)	(X)	(X)		(X)	(X)	
13b		Tank		(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)		(X)	(X)	(X)		(X)	(X)	
13c		Tank		(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)		(X)	(X)	(X)		(X)	(X)	
13d		Tank		(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)	(X)		(X)	(X)	(X)		(X)	(X)	
13e		Tank		X	X	X	X	X	X	(X)	(X)	(X)	X	X	X	(X)		X	X	X		X	X	(X)
14	Thickener Overflow	Weir	X			X	X		X	(X)	(X)	(X)	X	X	X	(X)								
15	Thickener Underflow Slurry	Pump Discharge																(X)	(X)	(X)				
16	Soda Ash Solution	Pump Discharge	X						(X)															
20	Filter Cake	Dock Blade			X													X	X	X		X	X	X
30	Limestone Feed (Dry)	Weigh Belt																		X	X	X	X	
31	Soda Ash (Dry)	Weigh Belt																		X	X			
40	Make-Up Well Water									(X)			(X)	(X)	(X)		(X)							
41	Make-Up River Water									(X)			(X)	(X)	(X)		(X)							

*In general, procedures in parentheses may not be required and will be performed on an as-needed basis; analyses not in parentheses will be performed on each sample unless experience suggests it is unnecessary.

To avoid changes in phase composition, if a phase separation is required to analyze a sample, it will be performed immediately after the sample is taken. These samples will only be taken while an Arthur D. Little technican or chemist is on duty and available to perform the separation.

3. Auxiliary Data

Additional information, not directly available from process instrumentation or sample analyses, may be necessary or desirable for a complete characterization of the process operation. A list of this information, along with the frequency of acquisition and the associated testing requirement is given in Table IV-7.

Of particular importance will be the fuel analyses. It will be necessary to have an analysis of the coal being fired during the test program to characterize the flue gas and also to provide backup data on the sulfur levels in the flue gas. An analysis will also be required for the fuel oil, both on ash levels and sulfur content, to correct outlet loadings for the contributions by the reheat systems. It is expected that only a few of these analyses will be performed during the test program.

A number of flue gas measurements will also be required during different periods of the test program, including occasional inlet and outlet water concentration and particulate sampling during particulate tests. The flue gas water levels may be required to check the assumption of adiabatic saturation in determining water evaporation and computing outlet flue gas rates which will be the basis for correcting SO_2 and particulate measurements across the reheater.

E. DATA REDUCTION AND ANALYSIS

General guidelines and procedures for the reduction and analyses of data are presented in this section. Typically, periods of a minimum of 24 hours in duration to as long as several weeks will be used for data analysis relative to overall system performance. Development of specific correlations to characterize the performance of a process unit will be based directly on the data recorded as opposed to the time averaged data for the 24-hour system performance. Summaries for the overall performance of the system throughout a given test period will also be developed. These will allow a comparison of the behavior of the system under different conditions. Furthermore, they will provide a check on material balances since some independent data may be used in these summaries. Whenever possible, the guidelines and procedures described below have been translated into equations which are presented in Table IV-8.

TABLE IV-7

AUXILIARY DATA

Data	Frequency	Remarks
Flue Gas Conditions:		
Particle Loadings on Venturi Inlet Gas/ Absorber Outlet Gas	As required	Only during particulate testing
Water Levels in Venturi Inlet Gas/ Absorber Outlet Gas	As required	If necessary to check evaporation rates and develop humidity correlations
Boiler Operation:		
Steam Rates	Summary of daily operation	Confirmation of inlet gas conditions
Coal Firing Rates		
O_2 Level in Flue Gas		
SO_2 in Inlet Gas (Lear-Ziegler Analyzer)		
Coal Analyses - S, HHV, H_2O, ash	Daily	
- $C\ell^-$	Biweekly	
Fuel Oil Analysis - S, HHV, ash, H_2O	Monthly (or shipment)	To adjust outlet flue gas conditions
Ambient Meteorological Data:		
Rainfall/Snowfall	Daily Summary	
Raw Materials:		
Limestone Shipments and Silo Inventories		Inventory data on raw materials received at the plant and charged to the system are to be maintained
Na_2CO_3 Shipments and Silo Inventories		
SO_2 Shipments and Inventories		

TABLE IV-8

PROCEDURES FOR REDUCING AND ANALYZING THE LIMESTONE DUAL ALKALI DATA

Procedures	Comments
1. Gas Flow $SCFM = \sum_i (\text{duration}) \times \frac{(SCFM)\ Calib}{(AMPS)\ Calib} \times (\text{Fan Amps})_i$	- Dry SCFM's
2. Process Streams Flow $GPM_i = \frac{\sum_i (GPM)_j\ [Na^+]_j}{[Na^+]_i}$	- i, stream under consideration j, streams of known flow rate and Na^+ concentration
3. SO_2 Loadings Inlet SO_2 ——— Direct reading Outlet $SO_2 = \frac{SO_2\ \text{Analyzer} \times (\text{System SCFM} + \text{Heating Gas}) - SO_2 \text{ by Fuel Oil (avg/minute)}}{\text{System SCFM}}$	- Check with Scholz's Lear-Ziegler Analyzer - Correct for reheat gas and S in fuel oil - SCFM in dry basis
4. SO_2 Removal $\Delta SO_2 = CF \sum_i (\text{duration})_i \times (SCFM)_i \times \left(SO_{2\ in} - SO_{2\ out}\right)_i$	- CF, conversion factor ppm to lb mole or lb
5. Soda Ash Feed a) S.A. = Δ Weigh Feeder Integrator Reading b) $S.A. = \sum_i (\text{duration})_i \times (\text{Feedrate})_i$ c) S.A. = Deliveries + Δ Silo Inventory	- Use to check material balance for overall period performance
6. Sodium Carbonate Consumption a) Gross Na_2CO_3 = S.A. x wt.% Na_2CO_3 in S.A./100 $\text{Gross Stoichiometry} = \frac{\text{Gross } Na_2CO_3 \text{ (in moles)}}{\Delta\ SO_2 \text{ (in moles)}}$ b) Net Na_2CO_3 = Gross Na_2CO_3 - System Volume x 1/2 $\left([Na^+]_{end} - [Na^+]_{start}\right)$ $\text{Net Stoichiometry} = \frac{\text{Net } Na_2CO_3 \text{ (in moles)}}{\Delta\ SO_2 \text{ (in moles)}}$	- For internal consistency use equivalent moles of Na_2^{++}
7. Sodium Loss in Cake a) $\text{Na Loss} = \frac{1/2\ Na^+ \text{ (moles/g dry cake) x cake rate}}{\Delta\ SO_2}$ b) $\text{Na Loss} = \frac{1/2\ Na^+ \text{ (moles/g dry cake)}}{SO_x \text{ (moles/g dry cake)}}$	- Na loss is given as moles % of Δ SO_2, where Na is expressed as Na_2^{++} - Caution, see text before using this method
8. Sodium Material Balance Net Na_2CO_3 = Na loss in cake + leaks + demister losses	- Expressed in $Na_2^{++}/\Delta\ SO_2$, mole %
9. Limestone Feed a) L.S. = Δ Weigh Feeder Integrator Reading b) $L.S. = \sum_i (\text{duration})_i \times (\text{feedrate})_i$ c) L.S. = Deliveries + Δ Silo Inventory	- Use to check material balance for overall period performance

TABLE IV-8
Continued

PROCEDURES FOR REDUCING AND ANALYZING
THE LIMESTONE DUAL ALKALI DATA

Procedures	Comments
10. Calcium Carbonate Consumption	
$CaCO_3$ = L.S. x wt.% $CaCO_3$/100	
Stoichiometry = $\frac{CaCO_3 \text{ (in moles)}}{\Delta\ SO_2 \text{ (in moles)}}$	
11. Calcium Content of the Cake	- Measured analytically
12. Calcium Material Balance	
$CaCO_3$ Consumption = Cake Rate x Calcium Content of Cake	- See 14 for cake rate
13. Limestone Efficiency	
L.S. EFF = $\frac{Ca\ SO_x \text{ in cake}}{\text{Total Ca in cake}} \times 100$	- Expressed as mole %
where Ca SO_x = Total Ca - $CaCO_3$	
14. Cake Generation	
a) Cake = Δ Weigh Conveyor Belt Integrator Reading	
b) Cake = $\frac{\Delta\ SO_2 \text{ (moles)}}{SO_x \text{ in cake (moles/g dry cake)}}$	- Caution, see text before using this method
15. Process Makeup Water	
a) Well Water = Filter Cake Wash + Vacuum Pump + Pump Seals	- Pump seal water requirement is estimated
b) River Water = Thickener Hold Tank + Soda Ash Solution + Limestone Slurry (Reactor T-1)	
16. Wash Ratio	
W.R. = $\frac{\text{Wash Water}}{\text{Cake Generated (Wet) x Cake Moisture}}$	- Expressed on a volume basis
17. Process Evaporation	
a) EVAP = Gas SCFM x Δ Water Content in Gas (out-in)	
b) EVAP = Feed-Forward to Absorber/Scrubber - Scrubber Bleed	
18. Water Balance	
Well Water + River Water + Rain = Absorber Evaporation + Open Tank Evaporation + Water Content in Cake + Δ Surge Tank Inventory (End-Start)	
19. Oxidation in Scrubber/Absorber	
$OXI = \left[(GPM \times [SO_4^=])_{\text{scrubber bleed}} - (GPM \times [SO_4^=])_{\text{feed forward}} \right] \times 100/\Delta\ SO_2$	- Expressed as mole % of Δ SO_2
where $GPM_{sb} = GPM_{ff} \times [\text{Total } Na^+]_{ff}/[\text{Total } Na^+]_{sb}$	

1. Flue gas Flow

The ducting configuration of the system is very restrictive for the installation gas flow rate monitoring equipment. Thus, during the previous lime dual alkali test program, fan amperage was used as a measure of gas flow rate based upon correlations developed from velocity traverses. The same procedure of using calibrated curves of volumetric flow rate vs. fan current will be used in this program. The existing correlations will be checked at the beginning of the program by pitot traverses.

2. Process Flow Measurements

The flow of some process streams, including makeup water, fluctuate because they are controlled by various liquid levels or on pH. For the purposes of material balances around specific process sections during relatively stable periods, the average flow rates through the system will be based upon the flow in two streams that are preset and maintained on automatic flow control: the fresh absorbent feed to the tray tower and the sodium carbonate makeup to the thickener. Total sodium ion concentration in the various streams will be used as a tracer material to relate the flows. This approach was successfully used in the previous lime-based prototype study at Scholz.

3. SO_2

SO_2 levels in both the inlet and outlet flue gas will be monitored separately by DuPont-400 analyzers. The samples will be drawn through specially designed conditioning systems, which essentially remove all the moisture in the sample.

The SO_2 level recorded by the inlet monitor reflects the sulfur released by the coal as well as any additional SO_2 injected upstream of the booster fan. The actual amount of sulfur released by the coal will be estimated by correcting for the amount of SO_2 injected. The SO_2 level recorded by the stack outlet SO_2 monitor reflects the SO_2 levels in the gas leaving the absorber as well as the contributions of sulfur and gas by the oil-fired reheater. The required correction to determine the actual SO_2 levels at the absorber outlet will be estimated from the amount of No. 2 fuel oil consumed, its sulfur content, and the amount of reheat gas. The flue gas leaving the system--expressed on a dry basis--is not the same as that entering the system since SO_2 and chlorine have been removed, and more importantly reheat gas has been added. Appropriate corrections will be made to determine the actual amount of flue gas leaving the absorber.

The amount of SO_2 removed is simply the difference in SO_2 loading--ppm by volume, dry basis, corrected to standard conditions--between inlet and outlet of the FGD system times the volumetric flow of flue gas expressed also on a dry basis and corrected to standard conditions.

4. Sodium

a. Soda Ash Feed

The amount of dry soda ash added to the system over short periods can be estimated in two ways:

- By taking direct daily readings from the weigh feeder integrator.
- By averaging the feed rate of soda ash using the length of time in which the rate remains relatively constant as the weight factor in the averaging procedure.

A third, more accurate method will be available to measure the soda ash consumption over longer periods of time such as during baseline testing or for test periods during the variational testing phase. For longer periods, the amount of soda ash fed to the system will be equivalent to the deliveries of soda ash during the period plus any change in silo inventory. Silo inventories will be determined by measuring the solids level at regular intervals (weekly) during the test program.

b. Sodium Carbonate Consumption

The purity of soda ash (sodium carbonate content) will be determined periodically to establish the actual amount of sodium carbonate fed to the system.

The net sodium carbonate consumption during a given period will be determined by correcting the total sodium carbonate fed to the system for any changes in sodium inventory in the system liquor.

The changes in sodium inventory will be determined by measuring the total sodium concentration in the clarified thickener liquor. As indicated in the sample frequency and analysis charts (Tables IV-5 and IV-6), the thickener overflow samples will be collected daily but only two to three samples a week will be analyzed unless experience dictates the need for more frequent analyses.

Often times, the sodium carbonate consumption is expressed in terms of sodium stoichiometry, which is defined as the molar ratio of sodium carbonate consumption to SO_2 removal by the system in only one period of time.

c. Sodium Loss in Cake

There will be two methods to estimate the sodium loss in the cake. Both methods will require the determination of total sodium in the cake (moles of Na^+/g cake). The methods differ in the way in which the cake generation will be estimated.

For the first method, the readings off the weigh conveyor belt at the discharge of the filter will be used to estimate the amount of cake produced.

The second method is based on the total sulfur content of the cake (moles of SO_x/g cake). The amount of cake generated will then be determined by simply dividing the amount of SO_2 removed from the flue gas by the SO_x content of the cake.

Care must be exercised in the interpretation of the sodium loss data since the solids being discharged with the cake will probably have been generated at least 12 hours before when the system might have been operating under a different set of conditions, particularly SO_2 removal and soda ash makeup rate.

d. Sodium Material Balances

The accuracy of sodium-related data will be checked by overall sodium material balances. Typically material balances will be performed for each test period; however, balances over shorter periods of time will be performed when necessary. Basically, the material balance consists of comparing, on a molar basis, the net sodium carbonate consumption against the sodium losses in the cake and any other losses such as leaks in pumps or pipes, demister losses, etc.

5. Limestone

a. Solid Limestone Feed

Same procedure as for solid soda ash feed (See 4a).

b. Calcium Carbonate Consumption

The calcium carbonate consumption is equal to the limestone fed, over the period of interest, corrected for the carbonate purity in the limestone. As in the case of sodium carbonate, the calcium carbonate consumption may also be expressed in terms of stoichiometry relative to the SO_2 removed.

c. Calcium Content of the Cake

The calcium content of the cake, both as $CaSO_3 \cdot yH_2O$ and unreacted $CaCO_3$, will be determined analytically by measuring the moles of calcium present in a gram of cake.

d. Calcium Carbonate Material Balances

On a molar basis, the amount of calcium carbonate fed to the system should be equal to the calcium content of the cake times the amount of cake generated in the period of interest. Inventory adjustments for accumulation/depletion of calcium in the process liquor should be negligible. Adjustments for accumulation or depletion of solids in the thickener, if necessary, will be based upon measurements of the depth of solids in the thickener. For the most part, calcium balances will be performed over sufficiently long periods to make such adjustments unimportant. Methods for measuring rates of cake generation are described below.

e. Limestone Efficiency

The limestone efficiency is defined as the ratio of $CaCO_3$ used by the system to regenerate the scrubbing solution to the total $CaCO_3$ fed to the system. It will be calculated by dividing the $CaSO_x$ content by the total available calcium content of the cake (the level of calcium in the cake adjusted for the $CaCO_3$/Ca ratio in the limestone). The $CaSO_x$ content of the cake will be estimated from analyses of total SO_4 and TOS in the cake corrected for sodium sulfur salts based upon the sodium content of the cake and the sulfur/sodium ratio in the mother liquor.

6. Cake Generation

The amount of cake generated over any period of time will be determined from the conveyor weigh belt integrator at the discharge of the vacuum filter. A second method is available as a check; the molar SO_x content of the cake is determined at certain intervals of time, and it is multiplied by the moles of SO_2 removed at the corresponding interval. The sum of these products over the period of interest yield the total cake generated. Care must be exercised in using the second method since the SO_x content of the cake might correspond to a different level of SO_2 removal due to the relatively long residence time of the solids in the system.

7. Water Process Makeup

Two types of process makeup water will be used in the system, river water and well water. Feed totalizers measure the total usage of water from each of these sources. The well water totalizer, however, also monitors the water feed to the pump seals; and no provisions exist in the system to either measure the pump seal leakage rates or to return the seal water leakage to the system. To the extent possible and practical, leakage from the pump seals will be checked to correct the well water flow totalizer readings.

a. Cake Wash Ratio

The wash ratio is defined as the amount of wash water used per unit volume of liquor contained in the waste cake. The amount of wash water will be obtained from a totalizer in the wash water stream. The water in the cake will be obtained by multiplying the cake generated (See 6) by the moisture content of the cake, which is determined analytically.

b. Scrubber Evaporation

The amount of water evaporated in the scrubber/absorber section to quench and saturate the flue gas may be estimated in two ways. The first method is based on simply measuring the water content in the inlet and outlet flue gas. These measurements, however, will be performed only infrequently; and, therefore, the second method will be routinely used, leaving the first one as a check. The second method simply accounts for the difference in volumetric flows of solutions in and out of the absorber/scrubber section. The difference could be estimated either from averaged flow rates or from the absorber feed flow and the increase in total sodium concentration across the scrubber and absorber.

c. Water Balance

The water balance is a simple relationship of inputs to the system (see above) and outputs: absorber and open tank evaporation and water content in the cake. The difference in inventory of the surge tank must also be accounted for.

8. Oxidation in the Scrubber

The total moles of $SO_3^=$ oxidized to $SO_4^=$ in the scrubbing section are determined by taking the difference between the moles of $SO_4^=$ being bled to the reactors and the moles of $SO_4^=$ in the feed forward to the absorber/scrubber. This oxidation may be compared to the total moles of SO_2 removed by the system and expressed on a percentage basis.

V. ECONOMIC CONSIDERATIONS

A. COSTS OF MODIFYING THE SCHOLZ PROTOTYPE SYSTEM

The modification and conversion of the 20 Mw lime-based dual alkali system for operation with limestone is being performed under Phases I and II of the project. The cost of modifying the system is being shared by EPA and CEA. The overall projected cost for the modification of the system is $613,500. The fraction of the total cost to be reimbursed by EPA amounts to $366,300; CEA's contribution is $247,200. A detailed breakdown of these costs is given in Table V-1.

B. GENERALIZED COST ESTIMATES FOR A NEW 500 Mw SYSTEM

A generalized estimate for the cost of installing and operating a limestone dual alkali system on a new 500 Mw boiler is presented in this section.

In general, a dual alkali flue gas desulfurization system consists of several sub-systems for the performance of specialized functions: absorption of SO_2 from the flue gas; regeneration of the spent scrubbing solution and precipitation of waste solids; separation of the regenerated solution from the solid waste generated; storage and preparation of raw materials; disposal of the waste; and provision of utilities. Basically, the system under consideration is assumed to be designed for a 95% SO_2 removal efficiency when burning coal containing 3.5% sulfur. Additional design basis for the system is detailed in Table V-2.

Although the dual alkali process is capable of removing particulate matter, this capability is not incorporated in the system under consideration. Thus, the absorbers are assumed to take flue gas, through booster fans, from high efficiency electrostatic precipitators. The booster fans for the dual alkali system also serve as induced draft fans for the boiler. The system is modular in nature with four absorber/two reactor trains. Each absorber is sized to handle 135 Mw. Consequently, there is a 10% spare capacity at full load; and the system is capable of treating at least 80% of the design flue gas load with one absorber down. Similarly the reactor system is capable of handling a full system load with one reactor train out of operation for short periods of time. The vacuum filters are operated and controlled independently and the number of filters in operation depends on the system load. Three filters are included--two operating and one full spare filter when operating at design conditions.

The dual alkali system is assumed to be installed in a new boiler located in the Midwest. The base year for cost estimates is mid-1979.

TABLE V-1

PROJECTED COSTS FOR MODIFICATION OF THE DUAL ALKALI SYSTEM

SCHOLZ STEAM PLANT

($000)

	Cost
Reimbursable Items	
• Prepurchased Parts:	
Modification to process	114.7
• Subcontract Items:	
Recommissioning	38.0
• Direct Labor:	
3642 Man-hours	51.0
• Labor Overhead:	41.3
• Travel & Per Diem:	5.1
• Consultants:	
Arthur D. Little	86.0
TOTAL DIRECT COST	336.1
• G&A:	
9%	30.2
TOTAL REIMBURSABLE COST	366.3
Non-Reimbursable Items	
• Prepurchased Parts:	
SO_2 Injection	5.4
• Subcontract Items:	
Recommissioning	181.4
Modification to process	49.3
SO_2 Injection	11.1
TOTAL NON-REIMBURSABLE COST	247.2
TOTAL MODIFICATION COST	613.5

TABLE V-2

DESIGN BASIS FOR THE GENERALIZED DUAL ALKALI SYSTEM

Boiler Capacity, Mw	500
Boiler Location	Midwest
Boiler Status	New
Coal Heating Value, Btu/lb	12,000
Btu/kw-hr	9,000
Boiler Operating time, hrs/yr	7,000
Coal Burned, tons/yr	1,312,500
S in Coal, %	3.5
Cl in Coal, %	0.10
SO_2 in Flue Gas (94% of S in coal), lb/hr	24,675
SO_2 at Scrubber Inlet, lb/10^6 Btu	5.48
SO_2 at Stack, lb/10^6 Btu	0.274
SO_2 Removal	95%
lbs/hr	23,440
tons/yr	82,044
Gas Flow Rate at Scrubber Inlet, acfm	1,540,000
Temperature, °F Scrubber Inlet, acfm	310
Flue Gas Oxygen, %	6
Oxidation in Dual Alkali System, %	10
Boiler I.D. Fan and Scrubber F.D. Fan	Common
Bypass	Yes
Flue Gas Bypassed for Reheat, %	0
Reheat Provided by Bypass Flue Gas, °F	0
Base Year for Cost Estimation	1979
Scrubber/Reactor Trains	4/2
Spare Capacity	
Absorbers, %	10
Filters, %	50
Pumps, %	100
Instruments, Necessary for Operation, %	100
Turndown to % of Capacity	20
Sludge Disposal	Onsite
Available $CaCO_3$ in Raw Limestone, %	90%
Limestone Utilization, %	95%
Limestone Storage Pile, days	30
Soda Ash Silo Capacity, days	15

1. Capital Investment

The estimated total capital investment for a generalized 500 Mw limestone dual alkali system is $47.4 million, which is equivalent to $94.8/kw. This cost does not include waste cake processing or disposal. Waste processing and disposal is taken as an operating cost.

Details of this capital investment are given in Table V-3. The process equipment and other material costs, which amount of $16 million, are further detailed in Table V-4. These material costs are based on the actual costs of a 300 Mw lime dual alkali system installed in Unit No. 6 of Louisville Gas and Electric's Can Run Station[1], and on estimated costs by Tennessee Valley Authority (TVA)[2]. The difference in number of modules due to size (500 Mw vs. 300 Mw) and the difference in reactor configuration and in raw material feed units between the lime and limestone based dual alkali systems are taken into consideration.

Common fans are used for the dual alkali system and the boiler. One fan is provided for each of the four absorbers. The fan cost for the dual alkali system is the incremental cost over that for each induced draft fan required for the operation of the boilers. The incremental cost is equal to the difference in cost between a fan used for a pressure drop of 27 inches WG and one used for 15 inches WG. The absorbers are equipped with spray nozzles for presaturation of the gas and trays for promoting gas/liquid contact.

Each of the two reactor trains has four agitated reactor tanks. Each reactor overflows sequentially into the next reactor. The last reactor, which overflows to the thickener, is also equipped with transfer pumps to drain the reactor system.

The soda ash silo has a storage capacity of 15 days for operation at design conditions. It is equipped with vibrating bottoms, weighing and feeding mechanisms. The limestone is first conveyed to a storage pile and from the pile to a day bin. The limestone in the bin is first crushed and then processed by a wet ball mill to reduce its size, prior to being fed to the system.

The cost of erecting the system is estimated at 50% of the process equipment and additional material costs. The service, utilities, and miscellaneous costs, taken as 6% of all other direct costs, cover items such as maintenance shops, analytical laboratories, communication, security, and office related costs; walkways; landscaping, etc. The indirect costs are estimated in direct proportion to the total direct costs: engineering at 9%; construction field expense at 11%; contractor's fee at 5%; and contingency at 20%.

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[1] Van Ness, R.P., et al., "Full-Scale Dual-Alkali Demonstration at Louisville Gas and Electric Co.: Final Design and System Cost," EPA-600/7-79-221b, September 1979.

[2] Tomlinson, S.V., et al., "Definitive SO_x Control Process Evaluations: Limestone Double Alkali and Citrate FGD Processes," EPA-600/7-79-177, August 1979.

TABLE V-3

ESTIMATED CAPITAL INVESTMENT FOR A
500 Mw LIMESTONE DUAL ALKALI SYSTEM
(1979 Dollars)

	$1000
• Direct Costs	
Process Equipment and Material Costs	16,090
Erection Costs	8,050
Services, Utilities, and Miscellaneous	1,450
Total Direct	25,590
• Indirect Costs	
Engineering Design (9% of D.C.)	2,300
Construction Field Expense (11% of D.C.)	2,810
Contractor's Fee (5% of D.C.)	1,280
Contingency (20% of D.C.)	5,120
Total Indirect	11,510
• Allowance for Startup and Modification	3,700
• Interest During Construction	4,450
Total Depreciable	45,250
• Allowance for Spare Parts	320
• Land	740
• Working Capital	1,080
Total Investment	47,390

TABLE V-4

ESTIMATED COST OF PROCESS EQUIPMENT AND MATERIALS
FOR A 500 Mw LIMESTONE DUAL ALKALI SYSTEM
(1979 Dollars)

Major Pieces of Equipment	
Absorbers	$ 1,850,000
Agitators	250,000
Ball Mill and Crusher	430,000
Fans and Fluid Drives	880,000
Miscellaneous Handling and Dust Collecting	310,000
Pumps and Motors	1,290,000
Reheaters	280,000
Tanks and Silos	1,610,000
Thickener Shell and Mechanism	500,000
Vacuum Filters	1,480,000
Weigh Feeders and Conveyors	230,000
	$ 9,110,000
Additional Material Costs	
Buildings	230,000
Ducting	1,570,000
Electrical and Instrumentation	1,710,000
Foundations and Structural Steel	1,210,000
Land Improvements and Miscellaneous	230,000
Lining	650,000
Piping and Insulation	1,380,000
	$ 6,980,000
	$16,090,000

Based on TVA's estimates, the allowance for startup and modification is taken at 10% of total direct and indirect costs; and the interest during construction is taken at 12% of total direct and indirect costs.

The allowance for spare parts is estimated at 2% of the process equipment and additional materials. The cost of land given in Table V-3 includes the land needed for the flue gas desulfurization system and for waste disposal, which is estimated at 210 acres. It is assumed that the expansion of the disposal site will be made progressively with the operation of the plant so the cost of land development is treated as an operating cost. The working capital requirement is estimated based on TVA's guidelines: 3 weeks of raw materials, 7 weeks of other direct costs, and 7 weeks of overhead.

2. Operating Costs

The estimated annual operating costs for a 500 Mw system are given in Table V-5. The average labor is estimated at three operators per shift, three shifts per day. The cost of one supervisor and one analyst are included in the labor costs. The plant overhead is estimated at 50% of the operating and maintenance costs. Administrative overhead is estimated at 10% of operating and maintenance labor. Maintenance materials are estimated at 5% of direct capital investment. Capital charges, including interest, depreciation, return on investment, taxes, and insurance are taken at 14.6% of the total fixed plant investment.

The total annual operating costs (including capital charges) amounts to $15.7 million, equivalent to 4.5 mills/kwh or $191/ton SO_2 removed.

TABLE V-5

ESTIMATED ANNUAL OPERATING COSTS FOR A
500 Mw LIMESTONE DUAL ALKALI SYSTEM
95% SO_2 Removal
(1979 Dollars)

Direct Costs	Annual Quantity	Unit Cost $	Annual Cost $
Raw Materials			
Limestone	143,990 tons	10/ton	1,439,900
Soda Ash	5,440 tons	90/ton	489,600
			1,929,500
Utilities			
Electricity	41.1×10^6 kw-hr	0.03/kw-hr	1,233,000
Water	324×10^6 gals	0.12/1000 gal	38,900
Steam	400×10^6 lbs	2.0/1000 lbs	800,000
			2,071,900
Labor and Maintenance			
Operating Labor	3 shifts	105,000/shift	315,000
Supervision	2,080 hours	18/hour	37,400
Analysis	2,080 hours	14/hour	29,100
Maintenance	5% of direct investment		1,140,000
			1,521,500
Sludge Disposal	2.78×10^5 wet ton	10/wet ton	2,780,000
TOTAL DIRECT COST			8,302,900
Indirect Costs			
Capital Charges	14.6% of Depreciable Investment		6,606,500
Overhead			
Plant	50% of Labor and Maintenance		760,800
Administrative	10% of Operating Labor		31,500
TOTAL INDIRECT COST			7,398,800
TOTAL ANNUAL OPERATING COST			15,701,700
Mills/kwh			4.5
$/ton SO_2 Removed			191

Basis: Operating time 7000 hr/yr
3.5% S in coal
Reheat - 35F°

Average SO_2 Removal - 391 lbs/min
Onsite Solids Disposal by Trucking